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A TECHNO-ECONOMIC EVALUATION OF CENTRAX DEASHING IN DIRECT COAL LIQUEFACTION

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ABSTRACT

A technical evaluation of the CENTRAX deashing process under development by Sandwell Technologies has been accomplished by analyzing process results obtained from both laboratory and prototype equipment. Conceptual direct coal liquefaction plants have been designed incorporating the CENTRAX process, and costs of liquid products from these plants have been estimated using the MITRE Coal Liquefaction Cost Model. These costs have been compared to those obtained using the Kerr-McGee Critical Solvent Deashing (CSD) process as an alternative deashing system. Wilsonville liquefaction runs 244 and 250 using Illinois No. 6 coal have been used as the basis for the conceptual commercial plants.

Suggested Keywords: Coal Liquefaction, Deashing, Centrifugal Separation, Staged Liquefaction, Solids-Liquids Separation, Critical Solvent Deashing.

This report was prepared by the MITRE Corporation, McLean, Virginia, USA.

MITRE Report Number WP-87W00029. Authors: David Gray and Glen Tomlinson

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EXECUTIVE SUMMARY

CANMET, the research and development division of the Department of Energy, Mines and Resources of Canada, has supported Sandwell Technologies of Canada in the development of the CENTRAX process which uses sequential centrifugation and extraction to separate solids from coal derived and coprocessing liquids.

They have conducted laboratory experiments, designed a prototype CENTRAX unit with a 250mm bowl diameter and operated the unit with coal liquids obtained from the Wilsonville coal liquefaction pilot plant (DOE, Catalytic, 1984).

CANMET has contracted the MITRE corporation to perform an unbiased technical and economic assessment of the CENTRAX process by performing technical and economic evaluations of the CENTRAX deashing process in the context of conceptual commercial integrated two-stage liquefaction plants based on scale up of the pilot plant operations at Wilsonville. The performance of the CENTRAX unit was then compared to the Kerr-McGee Critical Solvent Deashing (CSD) technology currently in operation at Wilsonville. This process uses a solvent near or at its critical temperature to selectively extract non-distillable coal liquefaction products from ash and unconverted coal.

The CENTRAX process uses a series (usually three stages) of specially designed centrifuges to concentrate the ash and unconverted coal into a compacted cake (the underflow). After each successive stage of centrifugation, the compacted cake is extracted with a selected boiling range cut of process derived liquid and recentrifuged. The objective of this extraction is to displace the heavy coal liquid contained in the interstices between the solid particles with lighter extractant material which can then be vaporized and recovered.

The special centrifuge equipment contains a piston-scraper that forms the false floor of the centrifuge bowl. After the feed coal-liquid extractant slurry has been centrifuged for a certain period of time, the feed is terminated and the piston-scraper is hydraulically raised within the bowl and scrapes the compacted cake upwards over the solids barrier at the top and into the solids trough surrounding the bowl. The scraper is then lowered to the bottom of the bowl, and feed is restarted. The CENTRAX centrifuge is thus a semi-continuous process.

MITRE has had prior experience in the evaluation of deashing systems for coal liquefaction. (MITRE, El Sawy, 1983) and has developed an integrated coal liquefaction cost model to investigate the potential economic impact of process improvements, for example improved deasher performance, on the cost of liquid products from conceptual commercial coal liquefaction facilities. This model simulates the liquefaction of 30,000 tons per day of coal to produce hydrotreated liquids. MITRE has also developed a computer simulation model of the CENTRAX deashing system so that estimates of flows through the different stages could be established. This computer simulation uses Sandwell's experimental data to estimate the expected performance of commercial scale CENTRAX units operating in the three-stage counter-currrent mode. In their own preliminary economic evaluation of CENTRAX, Sandwell assumed a level of performance that resulted in a compacted dried cake containing 96 weight percent solids. In the MITRE analysis, a dried compacted cake containing 95.2 weight percent solids is predicted only if it is assumed that the unconverted coal (UC) in the overflow converts on recycle, that all extractant and non-distillable oils can be completely recovered during the extractant recovery process, and that there is no preferential precipitation of asphaltenes (A) and preasphaltenes (PA) during the CENTRAX processing sequence. This series of assumptions is what MITRE terms the IDEAL case. Test results from the CENTRAX prototype unit showed significant selective precipitation of A and PA during the centrifugation-extraction cycle.

If it is assumed that the UC in the overflow does not convert but builds up until steady-state accumulation is removed and the non-distillate oils (maltenes) remaining in the cake cannot be recovered during the drying process, then the dried compacted cake would contain 86 weight percent solids. These assumptions are what MITRE terms the OPTIMISTIC case. This case is still optimistic because it still assumes that no precipitation of A and PA occurs during CENTRAX processing and that 100 percent recovery of extractant can be achieved.

If it is now assumed that the A and PA does preferentially precipitate so that their concentration in the compacted cake is doubled compared to that of perfect mixing, and that 5 weight percent of extractant remains in the cake after drying, the dried cake would contain 78 weight percent solids. This performance level is considered by MITRE to be the REALISTIC case. The economic and technical implications of CENTRAX deashing for all three of these performance levels have been evaluated and compared to Kerr-McGee Critical Solvent Deashing (CSD).

Wilsonville pilot plant data have been scaled using the MITRE liquefaction model to simulate conceptual commercial liquefaction plants. Data from run 244B have been used as representative of operations and performance levels achieved for the Integrated Two-Stage Liquefaction (ITSL) configuration (DOE, Catalytic, 1984). Run 250D has been used as representative of performance in the Reconfigured ITSL mode (DOE, Catalytic, 1986). Run 250G represents data from the RITSL model using the concept of ash recycle. In this mode, the majority of the high pressure separator bottoms from the hydrotreater are recycled directly back to stage one without deashing. A slip stream of the hydrotreater effluent is vacuum distilled and deashed to remove the steady-state accumulation of ash and unconverted coal.

Table ES-1 summarizes the CENTRAX performance parameters for conceptual commercial liquefaction plants at all three levels of assumptions covered in the MITRE analysis. ITSL, RITSL, and RITSL ash recycle configurations are included. For the ash recycle case (run

TABLE ES-1

SUMMARY OF CENTRAX PERFORMANCE PARAMETERS FOR CONCEPTUAL COMMERCIAL LIQUEFACTION PLANTS
Based on Wilsonville Two-Stage Liquefaction Pilot Plant Results

PERFORMANCE LEVEL	IDEAL	OPTIMISTIC	REALISTIC	CSD
Material lost in underflow	Theoretical *A+PA	Theoretical*A+PA+M**	A+PA+M+5%oil	—
Disposition of UC in overflow	converts	concentrates	concentrates	--
Run 244B (ITSL)				
% Solids in REJECT	96.41	90.17	81.85	70.60
No. of CENTRAX units***	4132	4459	4459	-
Run 250D (RITSL)				
% Solids in REJECT	95.24	87.03	78.53	62.80
No. of CENTRAX units	3497	3770	3770	-
Run 250G (1:0.4) (RITSL/Ash Recycle)				
% Solids in REJECT	90.14	76.27	66.64	68.80
No. of CENTRAX UNITS	2097	2312	2312	-
Run 250G (1:0.7) (RITSL/Ash Recycle)				
% Solids in REJECT	96.26	89.49	81.49	68.80
No. of CENTRAX units	2656	2872	2872	-

* Theoretical A+PA implies no selective precepitation occurs.

** M = Maltenes or non-distillate oils

*** Note number of CENTRAX units is actual number of centrifuges.

250G), two feed to extractant ratios are analyzed for CENTRAX : 1 to 0.4 and 1 to 0.7. The weight percent solids in the dried compacted cake are shown together with the estimated number of full scale CENTRAX units.

Table ES-2 summarizes the economic impact of using CENTRAX deashing on the cost of liquid products from the conceptual commercial liquefaction plants. Four sets of results are shown for three performance levels of CENTRAX compared to CSD. Raw and hydrotreated product costs are shown for comparison. The hydrotreated products place the quality of the liquids on an equal basis from all plant configurations and are, thus, a better basis for comparison. Two deasher capital cost factors (DAF) are used for CENTRAX. The base factor (1.0) and 1.5 times the base factor. Sandwell's cost estimate for the CENTRAX units forms the basis of DAF=1.0. An independent cost estimate for comparable size centrifugal equipment obtained from Sharples Inc. forms the basis of DAF=1.5.

Table ES-2 shows that, for realistic assumptions of CENTRAX performance, savings in hydrotreated liquid product costs of approximately only one percent can be expected when using CENTRAX in place of CSD deashing for ITSL and RITSL configurations. When ash recycle is used (250G) in the RITSL configuration, CENTRAX offers no savings compared to CSD. At best, CENTRAX represents a stand-off when compared to CSD. This analysis has, of course, assumed that the CENTRAX system can be scaled and that the problems encountered in the prototype unit with respect to solids cross-over and bearing design can be overcome. It also assumes a rapid scraper cycle time (20 seconds) to ensure minimal flow disruption during feed flow.

TABLE ES-2

SUMMARY OF CENTRAX ECONOMIC IMPACT COMPARED TO CSD (Costs in Dollars Per Barrel)

Wilsonville/Performance Run No. / Level	CENTRAX IDEAL		CENTRAX OPTIMISTIC		CENTRAX REALISTIC		Kerr-McGee CSD	
	Raw	Hydro	Raw	Hydro	Raw	Hydro	Raw	Hydro
244B DAF 1.0	33.99	37.79	35.58	39.15	36.30	39.77	38.11	41.33
244B DAF 1.5	35.04	38.69	36.76	40.17	37.51	40.82		
250D DAF 1.0	31.94	33.62	33.52	34.99	34.30	35.63	36.25	37.22
250D DAF 1.5	32.81	34.41	34.51	35.88	35.32	36.55		
250G (1:0.4) DAF 1.0	31.67	34.03	32.72	34.94	33.72	35.81	33.19	35.33
250G (1:0.4) DAF 1.5	32.23	34.54	33.35	35.52	34.39	36.42		
250G (1:0.7) DAF 1.0	31.61	33.99	32.05	34.38	32.57	34.82	33.19	35.33
250G (1:0.7) DAF 1.5	32.27	34.59	32.77	35.03	33.31	35.49		

The question is whether development of a specialized, novel centrifugal separator, which can only operate in a semi-continuous mode, is necessary. Modern vertical scroll centrifuges are capable of operating at temperatures in the range 240-340°C where viscosities of coal liquids would be dramatically lower than at 180°C, the temperature used in the Sandwell prototype. Also, development of improved coal liquefaction configurations such as has been demonstrated at Wilsonville during run 250G have changed the processing requirements for deashing systems. The neutral economic impact of using the CENTRAX system does not necessarily mean that centrifuges have no role to play in coal liquefaction deashing systems. Modern vertical scroll units operating in a single stage configuration may well be economically and technically attractive for the RITSL ash recycle mode of operation.

1.0 INTRODUCTION, SUMMARY AND CONCLUSIONS

1.1 Background

Direct coal liquefaction involves the production of liquid products from coal by using recycle solvent and hydrogen at elevated temperatures and pressures. Coal ash and unconverted coal (UC) remain in the coal liquids as solids in the conversion process and must be separated from the product liquids. Process efficiency and economy are improved if the ash and UC can be removed with a minimal loss of liquid extract.

CANMET, the research and development division of the Department of Energy, Mines and Resources of Canada, has supported Sandwell Technologies of Canada in the development of the CENTRAX process which uses sequential centrifugation and extraction to separate solids from coal derived liquids.

Sandwell has conducted laboratory experiments, designed a prototype CENTRAX unit and operated the unit with coal liquids obtained from the Wilsonville coal liquefaction pilot plant (DOE, Catalytic, 1984). The CENTRAX prototype unit has a centrifuge bowl diameter of 250 mm and has the capacity to process feed up to 10 liters per minute.

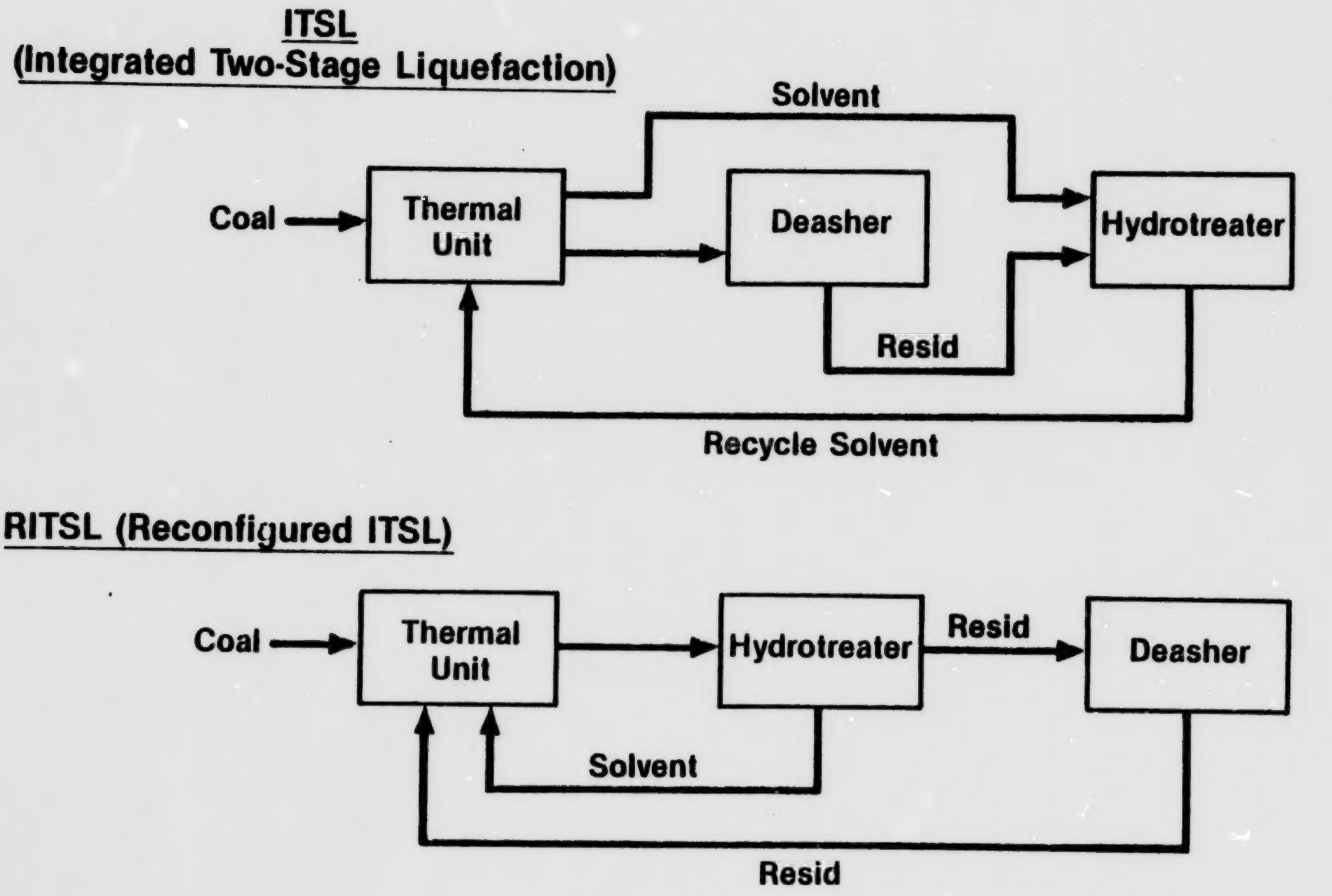
Sandwell has conducted an economic evaluation of their CENTRAX process in the context of a conceptual commercial two-stage coal liquefaction plant. Their analysis shows a substantial economic advantage over the Kerr-McGee Critical Solvent Deashing (CSD) process used at Wilsonville (Sandwell, Phase III 1985).

CANMET has contracted the MITRE corporation to perform an unbiased technical and economic assessment of the CENTRAX process. MITRE has had prior experience in the evaluation of deashing systems for coal liquefaction. (MITRE, El Sawy, 1983).

The Wilsonville pilot plant in Alabama has been investigating two-stage liquefaction using the configurations shown in Figure 1-1. In the Integrated Two-Stage Liquefaction (ITSL) mode, the thermally generated coal solution is deashed before the hydrotreater to prevent ash and unconverted coal from contacting the commercial hydrotreater catalyst. Ash and UC were once thought to contribute to drastic catalyst deactivation. However, subsequent work has shown that, ash and UC do not deactivate the catalyst when ebullated bed hydrotreaters are used.

Wilsonville has been investigating the Reconfigured Integrated Two-Stage Liquefaction (RITSL) mode where the thermally produced coal liquid containing the solid ash and UC is hydrotreated prior to deashing. This material has different characteristics from the thermally produced resid with respect to quantities of asphaltenes and preasphaltenes. The hydrotreated resid is generally lighter with fewer of these heavy asphaltic materials present and is, therefore, potentially an easier feedstock for deashing.

Wilsonville has concentrated on development of the Kerr-McGee Critical Solvent Deashing (CSD) process, (MITRE, El Sawy, 1983),. The CSD process extracts coal liquefaction products with an organic solvent near



ITSL (Integrated Two-Stage Liquefaction)

RITSL (Reconfigured ITSL)

Solvent
Coal
Thermal Unit
Deasher
Hydrotreater
Resid
Recycle Solvent
Coal
Thermal Unit
Hydrotreater
Resid
Deasher
Solvent
Resid

FIGURE 1-1
TWO-STAGE LIQUEFACTION CONFIGURATION

its critical temperature and pressure.

MITRE has developed an integrated coal liquefaction cost model to investigate the potential economic impact of process improvements on the cost of liquid products from conceptual commercial coal liquefaction facilities. This model simulates the liquefaction of 30,000 tons per day of moisture-free coal to produce hydrotreated liquids.

The model can predict the potential effect of improved deasher performance on liquid costs for a hypothetical deasher system that could achieve from 50 to 100 weight percent solids in the underflow. Figure 1-2 shows the results of such a prediction using the MITRE model. The base deasher cost is the cost of the baseline Kerr-McGee CSD system and the upper curve in Figure 1-2 shows the effect of increasing the deasher cost by 50 percent. The cost saving arises principally from increased recovery of coal liquids from the underflow and subsequent conversion of the material into distillate products. The cost model balances the conceptual commercial coal liquefaction plant for both hydrogen and fuel gas requirements. Thus, as additional residual material is recovered by improved deashing, less is available for gasification to produce hydrogen. Additional coal is used to satisfy this hydrogen requirement in the model. Figure 1-2 shows that, as percent solids is increased from 50 to 100 percent, product cost savings of 19 percent can potentially be realized.

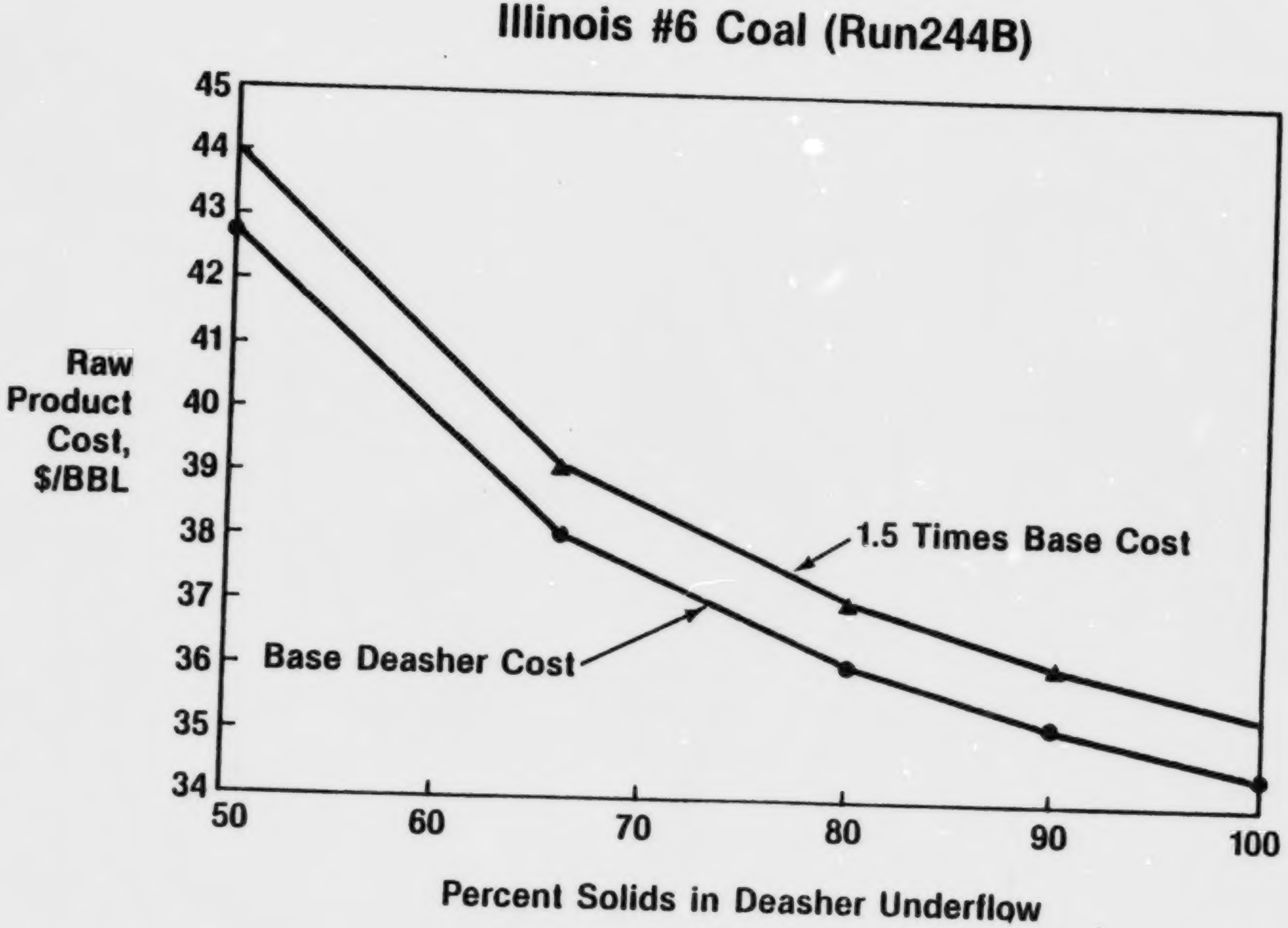


FIGURE 1-2
ECONOMIC EFFECT OF IMPROVED DEASHING

PROPRIETARY

Kerr-McGee CSD can achieve solids in the underflow in the range from 60 to 70 percent depending on feedstock and operating conditions. Sandwell Technologies (Sandwell) has claimed that the CENTRAX process can achieve levels of underflow solids in the range 95 to 100 percent. If CENTRAX can perform as claimed at a cost comparable to that of CSD, there could be a substantial savings in the cost of coal liquids produced.

1.2 Objectives and Methodology

The objectives of this study are as follows:

- technically assess the laboratory scale CENTRAX concept
- evaluate the experimental prototype testing program
- perform technical and economic evaluations of the CENTRAX deashing process in the context of conceptual commercial integrated two-stage liquefaction plants based on scale up of the pilot plant operations at Wilsonville.
- compare the economics of producing coal liquids from two-stage liquefaction using CENTRAX and Kerr-McGee Critical Solvent Deashing technologies
- offer recommendations to further development of the CENTRAX process

The methodology used to achieve these objectives is as follows:

- evaluate the technical performance of the CENTRAX unit by analyzing of laboratory scale and prototype results
- model a three-stage counter-current CENTRAX system to determine process flows in each stage
- scale the prototype unit to commercial size to determine commercial throughput per unit
- incorporate the CENTRAX model into the MITRE coal liquefaction cost model

- determine the cost of coal liquids using CENTRAX and compare to those using critical solvent deashing

1.3 Summary of CENTRAX Process Development

The CENTRAX process uses a series of specially designed centrifuges to concentrate the ash and UC into a compacted cake (the underflow). After each successive stage of centrifugation, the compacted cake is extracted with a selected boiling range cut of process-derived liquid and recentrifuged. The objective of this extraction is to displace the heavy coal liquid contained in the interstices between the solid particles with lighter extractant material which can then be vaporized and recovered.

The special centrifuge equipment contains a piston-scraper that forms the false floor of the centrifuge bowl. After the feed coal-liquid extractant slurry has been centrifuged for a certain period of time, the feed is terminated and the piston-scraper is hydraulically raised within the bowl and scrapes the compacted cake upwards over the solids barrier at the top and into the solids trough surrounding the bowl. The scraper is then lowered to the bottom of the bowl, and feed is restarted. The CENTRAX centrifuge is thus a semi-continuous process.

The Sandwell experimental program to demonstrate the CENTRAX concept consisted of laboratory and prototype testing (Sandwell, Phase I, 1985). The laboratory effort used a clinical laboratory centrifuge heated with hot air blowers to a temperature of about 180°C. Material to be separated was mixed with extractant, heated and transferred to the centrifuge where it was spun at 3200 rpm for five minutes. The centrifuge shield was removed, and the liquid decanted while hot and then allowed to cool. When necessary, several sequential extractions could be carried out using this procedure. The final process step, extractant recovery, was accomplished by using flash evaporation.

A series of single-stage extractions were performed to determine the effects of type and quantity of extractant, temperature, centrifugal force and residence time on the centrifuge performance. This performance was monitored by observing the ash concentration in the centrate or overflow, the solids concentration in the residue or compacted cake, and the amount of precipitation of asphaltenes and preasphaltenes.

Three-stage counter-current and cross-current configurations were also simulated by Sandwell. In the counter-current mode the overflow or separated liquid phase moves from stage three to stage one while the solids flow in the opposite direction. This allows clear extractant to contact the final compacted solids phase as a final clean-up step while the used separated liquids phase contacts the fresh feed.

Details of the procedures and results of these laboratory experiments are documented in Section 3 of this report. The laboratory-scale program demonstrated that heavy coal liquid materials from Wilsonville could be centrifuged with a process derived extractant in a heated laboratory centrifuge to produce a separated liquid phase containing between 0 and 0.2 wt. percent ash depending on conditions.

A limited number of operations in a semi-continuous mode were undertaken by Sandwell using the CENTRAX prototype equipment described earlier (Sandwell, Phase III, 1985). The process variables studied were feed flow rate, centrifugal force, feed type, and operating sequence of the scraper. During test runs, samples were taken of the separated liquid at various positions in the equipment, the compacted cake, and the feed. Simultaneous laboratory runs were performed with identical feeds to provide a comparison between the performance of the prototype and the laboratory centrifuge.

The detailed results of this investigation are documented in Section 4 of this report. Briefly, the results show that levels of unconverted coal or (UC) in the separated liquid are between 4 and 8 times greater than the level of ash for the prototype unit and about 20 times higher for the laboratory centrifuge. This relates to an inherent problem when using centrifugal devices for separating solids having large specific gravity differences, especially when the specific gravity difference is small between the liquid medium and one of the solids. The laboratory centrifuge also demonstrated 4 times higher ash removal efficiency than the prototype.

1.4 Economic Comparison of CENTRAX and Kerr-McGee CSD:

MITRE has analyzed the potential economic impacts of using CENTRAX deashing in place of CSD in the context of conceptual commercial coal liquefaction plants processing 30,000 tons per day of moisture-free coal to liquefaction.

In order to accomplish this, it was necessary to develop a computer simulation model of the CENTRAX deashing system so that estimates of flows through the different stages could be established. This iterative computer model simulated three-stage counter-current CENTRAX deashing. The MITRE model used Sandwell's experimental data for percent solids in the compacted cake, and the percent ash and UC recovered in the overflow as input.

In the simulation, the percent ash and UC were varied iteratively until the quantities in the stage three underflow agreed with the Sandwell test data. However, under the model assumptions of perfect liquid mixing and no preferential precipitation of heavier asphaltene (A) and preasphaltene (PA), the computer model predicted levels of A and PA over three times lower than achieved in the Sandwell tests. This suggests that there is significant selective precipitation of A and PA during the centrifugation-extraction cycle. It is estimated that final concentrations of A and PA in the underflow could be three times higher than would be predicted from perfect mixing assumptions.

Another significant observation is the preferential rejection of ash versus UC during centrifugation. If the UC that remains in the overflow is converted during subsequent recycle to the liquefaction reactors, then its retention is advantageous. However, if the UC remains refractory, it will build up so that the percent rejected equals that produced by the system on a steady-state basis.

This computer simulation developed by MITRE has been used to estimate the expected performance of commercial scale CENTRAX units operating in the three-stage counter-current mode. In their own preliminary economic evaluation of CENTRAX, Sandwell assumed a level of performance that resulted in a compacted dried cake containing 96 weight percent solids. In the MITRE analysis, a dried compacted cake containing 95.2 weight percent solids is predicted only if it is assumed that the UC in the overflow converts on recycle, that all extractant and non-distillable oils can be completely recovered during the extractant recovery process, and that there is no preferential precipitation of A and PA during the CENTRAX processing sequence. This series of assumptions is what MITRE terms the IDEAL case.

If it is assumed that the UC in the overflow does not convert but builds up until steady-state accumulation is removed and the non-distillate oils (maltenes) remaining in the cake cannot be recovered during the drying process, then the dried compacted cake would contain 86 weight percent solids. These assumptions are what MITRE terms the OPTIMISTIC case. This case is still optimistic because no precipitation of A and PA occurs during CENTRAX processing and 100 percent recovery of extractant is assumed.

If it is now assumed that the A and PA does preferentially precipitate so that their concentration in the compacted cake is doubled compared to that of perfect mixing, and that 5 weight percent of extractant remains in the cake after drying, the dried cake would contain 78 weight percent solids. This performance level is considered by MITRE to be the REALISTIC case. The economic implications of CENTRAX deashing for all three of these performance levels have been evaluated and compared to Kerr-McGee Critical Solvent Deashing.

Wilsonville pilot plant data have been scaled using the MITRE liquefaction model to simulate conceptual commercial liquefaction plants. Data from run 244B have been used as representative of operations and performance levels achieved for the Integrated Two-Stage Liquefaction (ITSL) configuration (DOE, Catalytic, 1984). Run 250D has been used as representative of performance in the Reconfigured ITSL mode (DOE, Catalytic, 1986). Run 250G represents data from the RITSL mode using the concept of ash recycle. In this mode, the majority of the high pressure separator bottoms from the hydrotreater are recycled directly back to stage one without deashing. A slip stream of the hydrotreater effluent is vacuum distilled and deashed to remove the steady-state accumulation of ash and unconverted coal.

Table 1-1 summarizes the CENTRAX performance parameters for conceptual commercial liquefaction plants at all three levels of assumptions covered in the MITRE analysis. ITSL, RITSL, and RITSL ash recycle configurations are included. For the ash recycle case (run 250G), two feed to extractant ratios are analyzed for CENTRAX : 1 to 0.4 and 1 to 0.7. The weight percent solids in the dried compacted cake are shown together with the estimated number of full scale CENTRAX units.

Table 1-2 summarizes the economic impact of using CENTRAX deashing on the cost of liquid products from the conceptual commercial liquefaction plants. Four sets of results are shown for three performance levels of

TABLE 1-1

SUMMARY OF CENTRAX PERFORMANCE PARAMETERS FOR CONCEPTUAL COMMERCIAL LIQUEFACTION PLANTS
Based on Wilsonville Two-Stage Liquefaction Pilot Plant Results

PERFORMANCE LEVEL	IDEAL	OPTIMISTIC	REALISTIC	CSD
Material lost in underflow	Theoretical *A+PA	Theoretical*A+PA+M**	A+PA+M+5%oil	—
Disposition of UC in overflow	converts	concentrates	concentrates	—
Run 244B (ITSL)				
% Solids in REJECT	96.41	90.17	81.85	70.60
No. of CENTRAX units***	4132	4459	4459	-
Run 250D (RITSL)				
% Solids in REJECT	95.24	87.03	78.53	62.80
No. of CENTRAX units	3497	3770	3770	-
Run 250G (1:0.4) (RITSL/Ash Recycle)				
% Solids in REJECT	90.14	76.27	66.64	68.80
No. of CENTRAX UNITS	2097	2312	2312	-
Run 250G (1:0.7) (RITSL/Ash Recycle)				
% Solids in REJECT	96.26	89.49	81.49	68.80
No. of CENTRAX units	2656	2872	2872	-

* Theoretical A+PA implies no selective precepitation occurs.
** M = Maltenes or non-distillate oils
*** Note number of CENTRAX units is actual number of centrifuges.

PROPRIETARY

TABLE 1-2

SUMMARY OF CENTRAX ECONOMIC IMPACT COMPARED TO CSD (Costs in Dollars Per Barrel)

Wilsonville/Performance Run No. / Level		CENTRAX IDEAL		CENTRAX OPTIMISTIC		CENTRAX REALISTIC		Kerr-McGee CSD	
		Raw	Hydro	Raw	Hydro	Raw	Hydro	Raw	Hydro
244B	DAF 1.0	33.99	37.79	35.58	39.15	36.30	39.77	38.11	41.33
244B	DAF 1.5	35.04	38.69	36.76	40.17	37.51	40.82		
250D	DAF 1.0	31.94	33.62	33.52	34.99	34.30	35.63	36.25	37.22
250D	DAF 1.5	32.81	34.41	34.51	35.88	35.32	36.55		
250G (1:0.4)	DAF 1.0	31.67	34.03	32.72	34.94	33.72	35.81	33.19	35.33
250G (1:0.4)	DAF 1.5	32.23	34.54	33.35	35.52	34.39	36.42		
250G (1:0.7)	DAF 1.0	31.61	33.99	32.05	34.38	32.57	34.82	33.19	35.33
250G (1:0.7)	DAF 1.5	32.27	34.59	32.77	35.03	33.31	35.49		

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CENTRAX compared to CSD. Raw and hydrotreated product costs are shown for comparison. The hydrotreated products place the quality of the liquids on an equal basis from all plant configurations and are, thus, a better basis for comparison. Two deashing cost factors (DAF) are used for CENTRAX. The base factor (1.0) and 1.5 times the base factor. Sandwell's cost estimate for the CENTRAX units forms the basis of DAF=1.0. An independent cost estimate for comparable size centrifugal equipment obtained from Sharples Inc. forms the basis of DAF=1.5.

Table 1-2 shows that, from realistic assumptions of CENTRAX performance, savings in hydrotreated liquid product costs of approximately only one percent can be expected when using CENTRAX in place of CSD deashing for ITSL and RITSL configurations. When ash recycle is used (250G) in the RITSL configuration, CENTRAX offers no savings compared to CSD. At best, CENTRAX represents a stand-off when compared to CSD. This analysis has, of course, assumed that the CENTRAX system can be scaled and that the problems encountered in the prototype unit with respect to solids cross-over and bearing design can be overcome. It also assumes a rapid scraper cycle time (20 seconds) to ensure minimal flow disruption during feed flow.

The question is whether development of a specialized, novel centrifugal separator, which can only operate in a semi-continuous mode, is necessary. Modern vertical scroll centrifuges are capable of operating at temperatures in the range 240-340°C where viscosities of coal liquids would be dramatically lower than at 180°C. Also, development of improved coal liquefaction configurations such as has been demonstrated at Wilsonville during run 250G have changed the processing requirements for deashing systems. The neutral economic impact of using the CENTRAX system does not necessarily mean that centrifuges have no role to play in coal liquefaction deashing systems. Modern vertical scroll units operating in a single stage configuration may well be economically and technically attractive for the RITSL ash recycle mode of operation.

2.0 DESCRIPTION OF CENTRAX CONCEPT

2.1 Background

Sandwell Technologies was awarded a contract with the CANMET entitled "Evaluation of the Integration of the Sandwell CENTRAX Process into an Integrated Two-Stage Liquefaction Process." The work was to be accomplished in three distinct phases. Phase I was an investigation on the laboratory scale of the CENTRAX concept. The Phase I effort (Sandwell, Phase I, 1985) included the following:

- development of reproducible analytical techniques for characterization of feed and product streams
- development of a laboratory-scale experimental procedure to investigate the effects of process variables on the CENTRAX performance
- simulation of multi-stage separation/extraction procedures
- simulation of final product recovery

Phase II of the contract was for the design and construction of a prototype CENTRAX separation unit (Sandwell, Phase II, 1985). Phase III was for testing the CENTRAX concept using the prototype apparatus (Sandwell, Phase III, 1985).

Sandwell has completed all three phases of the contract and has produced three reports documenting the effort.

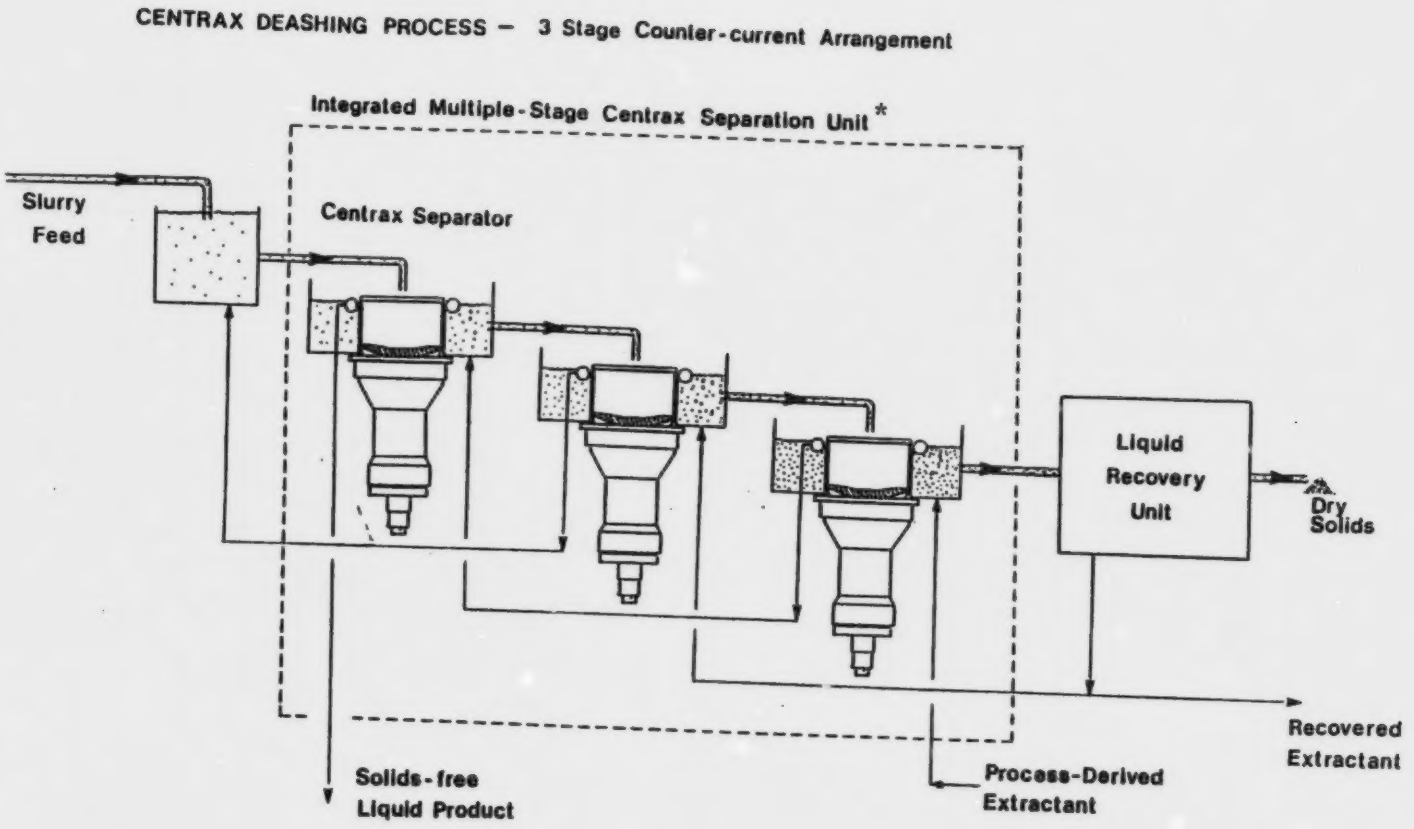
2.2 The CENTRAX Concept

The CENTRAX process is a technology designed to effect a complete separation of solid ash and unconverted coal material from coal derived liquids produced during the direct liquefaction of coal or the coprocessing of coal and residual oils. Depending on the process used, product liquids are a mixture of distillable and non-distillable oils, asphaltenes and preasphaltenes in varying proportions. Dispersed in this mixture are particles of coal ash and unconverted particles of organic coal material. Because the non-distillable oils (maltenes), asphaltenes and preasphaltenes cannot be separated by distillation and, for economic reasons they cannot all be rejected, then some technique must be used to recover these materials and obtain them free from solid contamination. Several techniques are potentially available to effect this recovery. Among them are filtration, centrifugation, Critical Solvent Deashing (CSD), and Anti-Solvent Deashing. Each one of these techniques has its own particular set of problems and disadvantages, but one disadvantage that is common to all is the rejection or loss of a portion of the product liquid. The amount rejected depends on the process, but generally filtration has the lowest loss, and anti-solvent deashing the greatest.

The CENTRAX concept is an attempt to eliminate this rejection of product liquid by recovering almost 100 percent of the organic material. The CENTRAX process, in theory, accomplishes this by the displacement of the heavy, high molecular weight coal liquid material including heavy oils, asphaltenes, and preasphaltenes from the solids using a process-derived extractant which is subsequently recoverable by some unspecified evaporative process. This extractant should also have the property of solubilizing the oils, asphaltenes, and preasphaltenes throughout all the process steps.

The CENTRAX process uses specially designed centrifuges in stages as depicted schematically in Figure 2-1 which shows three such stages. In the process the feed slurry containing the solids to be separated is fed into the Stage I separator where the ash and unconverted coal are separated from the coal liquids and extractant solution by application of centrifugal force. The extractant used in the first stage comes from the second stage separated liquids overflow as shown in Figure 2-1. After centrifugation in Stage I the separated liquid product, which is theoretically solids free exits the unit. The concentrated solids are mixed in the solids trough surrounding the centrifuge with separated liquids from Stage III and the resultant slurry is the feed to Stage II. After centrifugation in Stage II, the separated liquids are fed to the slurry feed tank and the separated solids are extracted with recovered extractant from the liquid recovery unit. The slurry is then feed to Stage III. After centrifugation, the separated liquids are used as the extractant in the Stage II feed and the separated solids are contacted with fresh process-derived extractant and then sent to the liquid recovery unit where, in the ideal world, the extractant is recovered and the solids are produced as a dry, friable material.

This counter-current configuration uses extractant to displace the product liquids (oils, preasphaltenes, and asphaltenes) at each stage of the separation/extraction process. In the final stage the concentrated solids are contacted with the fresh extractant material to displace the


CENTRAX DEASHING PROCESS – 3 Stage Counter-current Arrangement
Integrated Multiple-Stage Centrax Separation Unit *
Slurry Feed
Centrax Separator
Liquid Recovery Unit
Dry Solids
Recovered Extractant
Process-Derived Extractant
Solids-free Liquid Product

*Note: in this report numbers of CENTRAX units refers to the total number of centrifuges.

FIGURE 2-1
CENTRAX DEASHING PROCESS – 3 STAGE COUNTER-CURRENT ARRANGEMENT

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last of the interstitial product liquids. The extractant, being a ligher boiling material, is then flashed off to produce the dry, free flowing ash and unconverted coal solids mixture.

The centrifugal separation unit is the heart of the CENTRAX process and was designed and constructed especially for this task. The CENTRAX separator, described in Section 4 of this report, consists of two sections. The upper section is the centrifugal rotating bowl surrounded by the solids trough. This is the processing section. The lower part is the hydraulically operated piston-scraper assembly. In its lowest position, the scraper forms the false floor of the centrifuge bowl. The scraper is designed so that it has a 1 mm clearance between the end of the scraper blade and the bowl wall. The scraper assembly rotates at the same speed as the bowl, but because of the clearance, a layer of solids cake 1 mm thick builds up on the bowl wall. This layer prevents the scraper blade from eroding and contacting the bowl wall. After the slurry has been centrifuged for a certain time, the hydraulic mechanism is activated and the scraper moves upward. As the scraper is raised inside the bowl, it scrapes the deposited solids cake upward and over the solids barrier at the top of the bowl into the solids trough. During this operation the slurry feed to the unit is stopped. After the scraper reaches the top of the bowl and has removed all but a 1 mm layer of solids from the wall, the hydraulics are reversed and the piston-scraper assembly is lowered to the bottom of the bowl. The slurry feed is then restarted, and the cycle can begin once more.

The CENTRAX concept is thus a combination of certifugal separation and solvent extraction. If the low boiling solvent can replace the heavy product liquids that adhere to the solid particles and are retained in the interstitical void space, then this low bowling extractant should be recoverable by some form of evaporative or spray-drying technique. The prototype unit was used to demonstrate one stage of this separation/extraction process, and analyses were made of the separated liquids and concentrated solids. However, no attempt appears to have been made during the prototype development program to simulate multi-stage operations except at the laboratory scale. Also, apart from a laboratory scale evaporative technique, the concern of final extractant recovery was not addressed in this program.

3.0 LABORATORY SCALE DEVELOPMENT EFFORT

3.1 Experimental Procedure

The laboratory scale development effort undertaken by Sandwell (Sandwell, Phase I, 1985) attempted to simulate the three operations that are involved in the complete CENTRAX scheme. These operations are depicted in Figure 3-1. Step 1, preheating, was accomplished in the centrifuge shield itself (i.e., no insert tube was used), because a larger volume of material could be processed and the shield provided a wider bore opening for mixing and transfer processes. The centrifuge shield containing the feed and extractant were simply stirred with a spatula in a constant temperature oil-bath until it was assumed that homogeneity was reached. After preheating, the shield was transferred by gloved hand to a clinical laboratory centrifuge.

The centrifuge was heated with hot air blowers and the temperature was maintained and monitored by a thermocouple. After centrifugal separation, usually performed at 3200 rpm for five minutes, the centrifuge shield was decanted while hot and then allowed to cool to ambient temperature. If deemed necessary, several extractions could be carried out in sequence on the feed material. The final step, extractant recovery, was accomplished by what Sandwell terms "flash evaporation." The final version of this technique involved a heated inclined stainless steel plate enclosed in an air-tight steel box through which nitrogen was passed. The nitrogen swept the vapors from the plate into cooled traps for liquid collection. An injection system consisting of a stainless

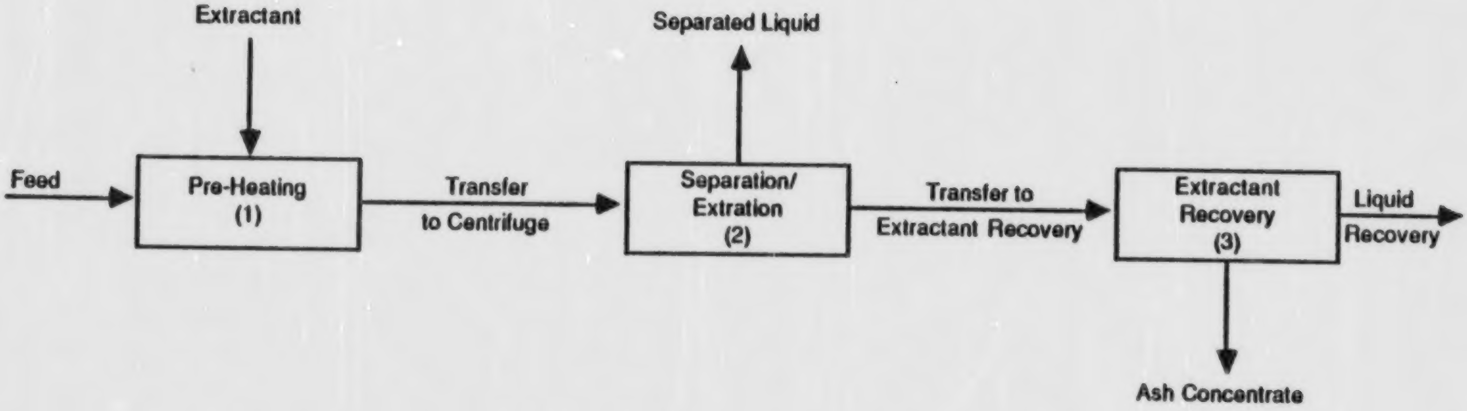
Extractant
Feed
Pre-Heating (1)
Transfer to Centrifuge
Separated Liquid
Separation/ Extration (2)
Transfer to Extractant Recovery
Extractant Recovery (3)
Liquid Recovery
Ash Concentrate

FIGURE 3-1
LABORATORY-SCALE PROCESS SIMULATION

PROPRIETARY INFORMATION

steel tube fitted with a hand operated piston pusher was used to introduce the concentrated solids. Sandwell apparently favored this system over vacuum evaporation because it was easier to operate and produced a friable ash-concentrate that was non-sticky and easy to handle. This extractant recovery system was only a laboratory expedient and was not intended for commercial scale-up.

3.2 Feedstock Characterization and Preparation

Coal liquefaction products from the Wilsonville pilot plant in Wilsonville Alabama were used in the investigation. Table 3-1 shows an analysis of the feedstocks. V110 BTM 243/242 are the flashed reactor effluents before vacuum distillation from runs 242 and 243 using Illinois Number 6 coal. T102 BTM 243/242 are the vacuum tower bottoms material that would normally be the feed to the Kerr-McGee CSD (Critical Solvent Deashing) unit. T105 BTM 243 is the hydrotreated solvent bottoms stream from the distillation tower in the hydrotreating unit. This material is 100% hexane soluble and has an initial boiling point of 180°C and a final boiling point of 509°C.

The solids containing materials are shiny black solids at room temperature. To achieve homogeneity, they were melted to 110°C and mixed thoroughly with a mechanical mixer. Each mixture was then cooled and flaked. The flakes were then frozen, ground, and mixed again. The samples were analyzed for oils (O), asphaltenes (A), preasphaltenes (PA), unconverted coal (UC) and ash. Vacuum distillations were performed to give the percentage of distillate products and the distillate fractions

TABLE 3-1
ANALYSIS RESULTS OF WILSONVILLE LIQUID COAL PRODUCTS AS RECEIVED

Process Stream	Components (wt %)				
	O	A	PA	UC	Ash
V110 BTM 243	46.20	29.77	14.38	5.12	4.53
V110 BTM 243	49.59	26.56	15.40	4.36	4.09
T102 BTM 243	12.70	48.94	25.22	7.60	5.54
T102 BTM 242	14.36	42.39	31.69	5.80	5.78
T105 BTM 243	100.00	-	-	-	-

were further analyzed by simulated distillation.

3.3 Single-Stage Separation/Extractions

A series of single-stage extraction experiments were performed as indicated in Section 3.1 to determine the effect of the following parameters on the performance of the centrifuge technique:

- type of extractant
- amount of extractant
- temperature of extraction
- centrifugal force
- residence time.

The performance was monitored by observations of the following dependent variables.

- ash concentration in the overflow (or separated liquid)
- amount of precipitation of A and PA
- compressibility of solid concentrates.

Figure 3-2 shows the results of measurements of the ash content of the separated liquids for the feed material V110 BTM 243 (containing 4.53 percent ash) using three different extractants at various feed to extractant ratios. The extractants used were:

- T105 BTM 243
- V110 BTM 243 distilled to an end point of 400°C
- V110 BTM 243 distilled to an end point of 500°C

Four feed to extractant ratios were used varying from 1:0.2 to 1:3. The single-stage experiments were conducted at a centrifugal acceleration of 1800 Gs at 180°C for five minutes.

These results show that extractant T105 BTM 243 was effective at all extractant levels to achieve a zero ash level in the overflow. The other

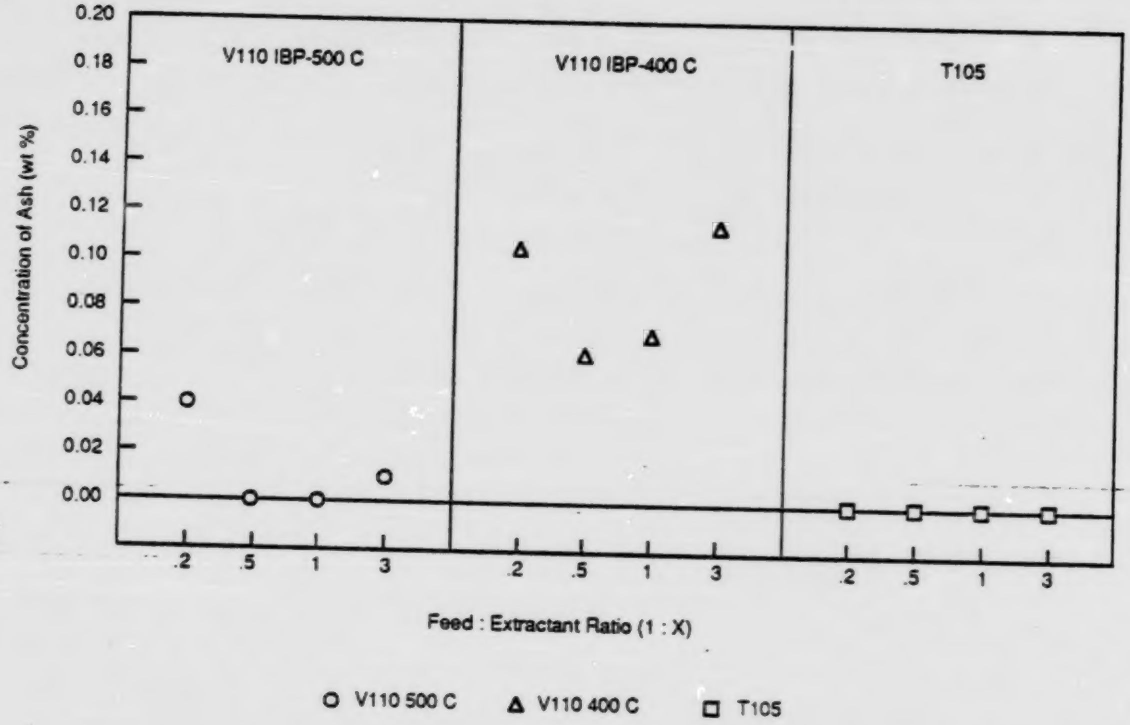

FIGURE 3-2
ASH CONTENT -SEPARATED LIQUID
(TYPE AND AMOUNT OF EXTRACTANT)

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extractants generally were not as effective. Figure 3-3 shows the ash concentration in the overflow for various feeds at different feed to extractant ratios. It can be seen that essentially no ash separation was achieved using the heavy T102 materials in the absence of extractant. However, low ash levels were obtained with the V110-242 material even without extractant. The V110-243 material required a feed to extractant ratio of at least 0.3 to achieve zero ash levels in the overflow. When this material was centrifuged at 130°C instead of 180°C, ash levels of 0.16 percent were still found in the overflow even at a feed to extractant ratio of 3:1. This shows the importance of viscosity dependency on temperature for this system.

The effects of residence time and centrifugal force were also briefly examined in the program. Table 3-2 shows that high G forces and long residence time were necessary to produce the lowest ash concentration in the overflow. This result is as would be expected from these systems.

Producing a clean, ash-free overflow is only one objective of solids/liquids separation. The recovery of product liquid material from the concentrated solids is of particular importance for the economics of coal liquefaction or coprocessing. The objective of the CENTRAX program is to recover essentially all of the solubilized coal or product material including asphaltenes and preasphaltenes. Thus, it is desirable to prevent, as far as possible, the preferential concentration of these materials with the solids, since, if this occurs, these materials will be rejected with the solids and thus lost for further upgrading into liquid products.

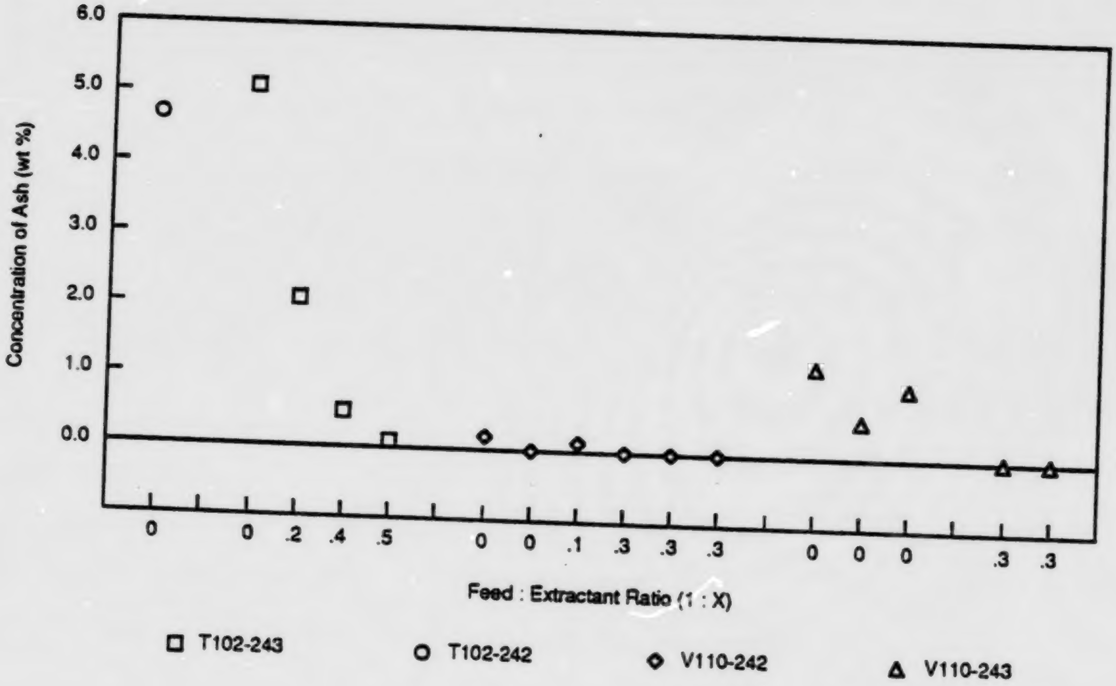


FIGURE 3-3
ASH CONTENT -SEPARATED LIQUID
(TYPE OF FEED AND AMOUNT OF EXTRACTANT)

PROPRIETARY INFORMATION

TABLE 3-2
ASH CONCENTRATION OF SEPARATED LIQUID FOR V110 BTM-243 AND T105 BTM 243 IN A RATIO OF 1:01

	Separation Residence Time	
	5 min	10 s
Centrifugal Force		
1800 G	0.18%	1.78%
700 G	1.24%	2.37%

Sandwell attempted to monitor the preferential concentration of A and PA by measuring the relative concentrations of these materials before and after centrifugation. A large effort appears to have been devoted to the exercise including determination of amounts of precipitation with various feeds and feed to extractant ratios for the single-stage centrifugation scheme. Variations of preferential separation with centrifugal force, temperature and residence time were also investigated. In spite of the large data base generated by these studies, Sandwell concedes that the difficulty in accurate measurement of tetrahydrofuran (THF), toluene, and hexane insolubles results in considerable scatter in the data.

In order to bring some level of consistency to the scattered data, Sandwell defined a Precipitation Index P as:

$$P = \frac{\text{(O/PA) in Feed (extractant and solids free basis)}}{\text{(O/PA) in Separated Liquid (extractant and solids free basis)}}$$

An index of 1 indicates no preferential separation of PA. P values greater than 1 measure the degree to which PA have been precipitated, i.e., essentially lost to the system as reject. Sandwell plotted P index against feed to extractant ratio for ratios of 0 to 3.0. The trend indicated that as the feed to extractant ratio increased so did the value of P. Thus the amount of PA precipitation increased with an increasing amount of extractant. This result seems to be at odds with the CENTRAX objective. However, the result is intuitively satisfying. Since preasphaltenes are more dense than oils and not likely to be completely soluble in the extractant, some selective precipitation under a force of

3000G is possible if not probable. The less viscous media resulting from use of more extractant would be expected to increase the PA separation, i.e, it may act to some degree as an anti-solvent. Multi-stage techniques described below can, however, increase the recovery of PA material even if some preferential precipitation should occur.

3.4 Multi-Stage Extractions

Sandwell simulated counter-current separation/extraction operations by using a three-stage cascade mode shown in Figure 3-4. The stages are numbered in the direction of solids flow. The SL (separated liquid) phase is the liquid overhead from each stage in a direction opposite to the solids flow. Thus clean extractant contacts the final CS (compacted solids) phase as a final clean-up step while the used SL phase contacts the fresh feed. In these experiments Sandwell used V110 BTM 243 and T105 BTM 243 as feed and extractant in the ratio 1:0.3 Experiments were all conducted at 180°C, 1800 Gs for five minutes. Four cycles were performed before steady state was achieved.

Figure 3-5 summarizes the mass balances obtained in this simulation. Sandwell claims that in this operation the ash concentration of the separated liquid phase was 0.08 percent which corresponds to an ash removal of 98.3 percent from the separated liquid. However the ash balance is only 94 percent based on input ash. Thus, a quantity of ash over three times that reported in the overflow has apparently been unaccounted for. The unconverted coal material balance was only 68 percent which indicates that either the process converts UC or they failed

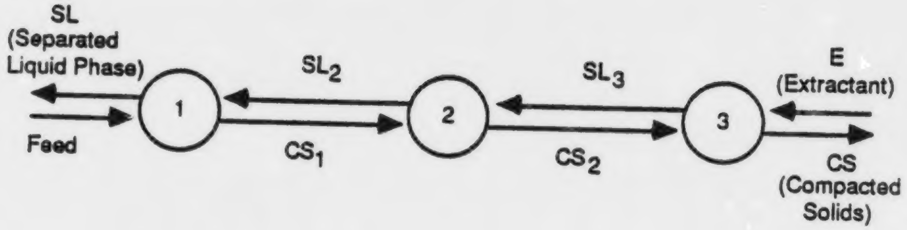
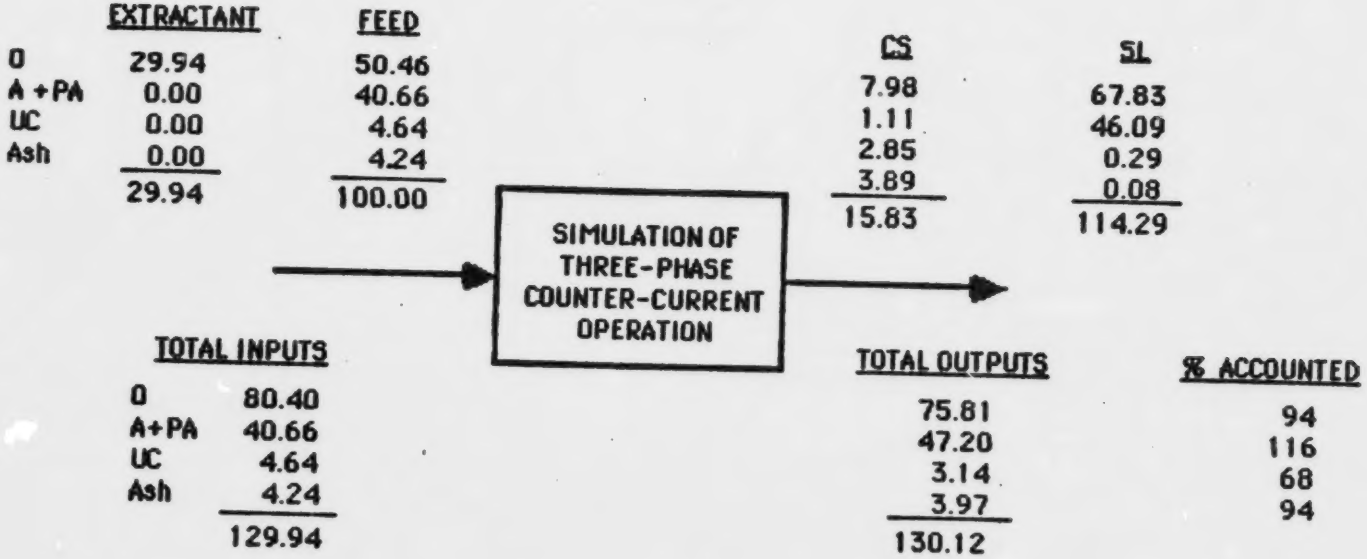

SL
(Separated
Liquid Phase)
Feed
1
SL2
CS1
2
SL3
CS2
3
E
(Extractant)
CS
(Compacted
Solids)

FIGURE 3-4
SCHEMATIC FLOW DIAGRAM OF THREE STAGE
COUNTER-CURRENT EXTRACTION

PROPRIETARY INFORMATION


SIMULATION OF
THREE-PHASE
COUNTER-CURRENT
OPERATION

	EXTRACTANT	FEED	CS	SL
O	29.94	50.46	7.98	67.83
A+PA	0.00	40.66	1.11	46.09
UC	0.00	4.64	2.85	0.29
Ash	0.00	4.24	3.89	0.08
	29.94	100.00	15.83	114.29

	TOTAL INPUTS	TOTAL OUTPUTS	% ACCOUNTED
O	80.40	75.81	94
A+PA	40.66	47.20	116
UC	4.64	3.14	68
Ash	4.24	3.97	94
	129.94	130.12	

FIGURE 3-5
SIMULATED COUNTER-CURRENT SEPARATION/EXTRACTION RESULTS

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to account for it in the overflow.

In addition to counter-current separation, Sandwell also performed experiments using simulated three-stage cross-current configurations. Figure 3-6 shows the principle of cross-current extraction/separation. In this scheme the solids phase is successively contacted with fresh extractant at each stage. Thus for three-stage operation, three separated liquid phases and a single concentrated solids phase result.

Figure 3-7 shows the mass balances obtained using this configuration. In this case the mass balance closure is very good, but the weighted average solids concentration in the overflow is high at 0.88 percent. Unfortunately, Sandwell does not differentiate UC and Ash in these results so the ash concentration cannot be determined. The CS contains 33.5 wt percent solids. A+PA recovery is good at 98 percent. Cross-current operations, however, require a large quantity of extractant; almost 1:1 on a weight basis compared to counter-current operations which have been using approximately 1:0.3.

3.5 Extractant Recovery

Following the extraction/separation centrifugal phase, the final stage of the process is the recovery of the extractant by some form of evaporative technique. Sandwell, as previously mentioned, uses a flash

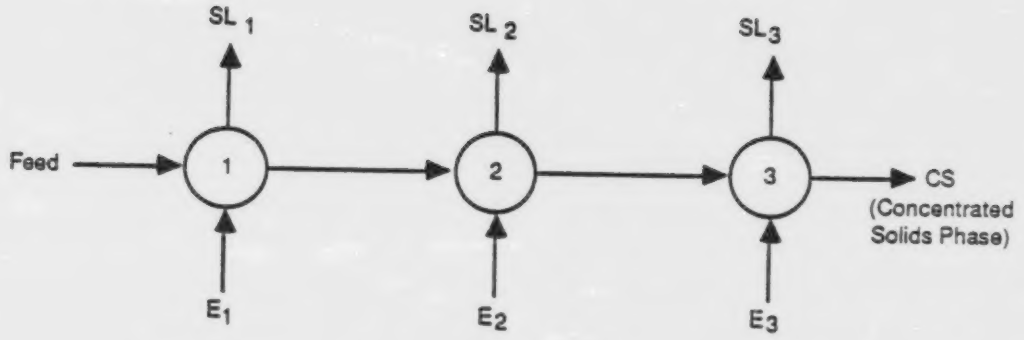
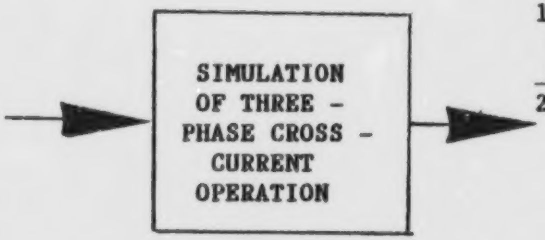

SL 1
SL 2
SL3
Feed
1
2
3
CS
(Concentrated
Solids Phase)
E1
E2
E3

FIGURE 3-6
SCHEMATIC FLOW DIAGRAM OF THREE-STAGE
CROSS-CURRENT EXTRACTION/SEPARTION

PROPRIETARY INFORMATION

	FEED	$E_1+E_2+E_3$
O	45.85	98.30
A+PA	44.87	0.00
UC	4.75	0.00
Ash	4.53	0.00
	100.00	98.30


SIMULATION OF THREE - PHASE CROSS - CURRENT OPERATION

CS	SL1	SL2	SL3
16.04	65.93	31.36	31.17
0.04	41.03	2.45	0.64
8.12	1.47	0.05	0.01
24.20	108.43	33.86	31.82

	Total Inputs	Total Outputs	% Accounted
O	144.15	144.50	100
A+PA	44.87	44.16	98
UC	9.28		104
Ash		9.65	
	198.30	198.31	

FIGURE 3-7
SIMULATED CROSS-CURRENT SEPARATION/EXTRACTION RESULTS

evaporation apparatus to simulate this recovery stage. Table 3-3 summarizes the results of trouble-free experiments for the feed materials V110 BTM 242 and V110 BTM 243. The product recovery (PR) is defined:

$$PR = \left(100 - \frac{\text{Total Solids in AC* - Total Solids in Feed}}{\text{(Total Feed - Total Solids in Feed)}}\right)$$

The results show that the residual solid is concentrated as a product which is 100% THF insoluble, dry and friable and that high product liquid recoveries are possible.

3.6 Observations

The laboratory scale development effort has shown that heavy coal liquid materials from Wilsonville can be centrifuged with a process derived extractant in a heated laboratory centrifuge unit to produce an overflow containing between 0 and 0.2 wt percent ash depending on conditions. Liquid products, A, and PA can be recovered from the ash concentrate by using successive extractions followed by centrifugation. Simulated extractant recovery using a hot inclined plate has been demonstrated.

The laboratory tests are obviously batch mode operations while commercially the process has to be demonstrated in the continuous mode. A limited number of operations in a semi-continuous mode has been undertaken in the CENTRAX prototype unit to be discussed in the next section.

---

*AC = Ash Concentrate

TABLE 3-3
EXTRACTANT RECOVERY FROM ASH CONCENTRATE - SUMMARY*

Starting Material	Mass Balance %	Gas Produced %	Coke Produced %	Overall Product Recovered %	THF Insol. Fraction of Ash Concentrate %	Description of Ash Concentrate %
V110 BTM 243 T105 BTM 243	95.7	4.3	0	100.2	100	Dry friable
V110 BTM 242 T105 BTM 242	100	0	10.5	89.5	100	"
V110 BTM 242 T105 BTM 243	94.6	0	100.0	100	"	

*Tests conducted at 550°C Plate Temperature.

4.0 CENTRAX PROTOTYPE TESTING AND EVALUATION

4.1 CENTRAX Pilot Plant Description

The CENTRAX pilot plant consists of four heated vessels with recirculation pumps, transfer lines, a feed buffer tank, and the CENTRAX centrifugal unit. One vessel contains the slurry feed, another contains the extractant, a third vessel is for collecting the separated liquids, and the fourth vessel is for collecting the separated solids.

The CENTRAX separator is shown schematically in Figure 4-1 (Sandwell, Phase III, 1985). The top part is the centrifugal bowl separator, and the bottom part is the hydraulic drive mechanism which operates the piston scraper.

During operation of the unit, pre-heated slurry is fed into the feed buffer tank and from there, through a control valve, into the centrifuge bowl. Flow into the buffer tank is continuous while flow to the centrifuge is controlled by an electrically operated ball shut-off valve. When closed, the slurry accumulates in the buffer tank until the valve is opened. Control of this valve can be automatic (i.e., linked to the position cycle of the solids scraper) or manual. When the valve is opened the pre-heated slurry stream is directed into the bowl onto the top of the solids scraper. This is turning at about 4600 RPM together with the hydraulic cylinder, piston, piston rod, and separator bowl. This spinning motion forces the slurry against the bowl wall. The solids, because of their higher density, form a cake on the bowl wall while the liquid is displaced into a layer on top of the cake. As the feed is continued, the

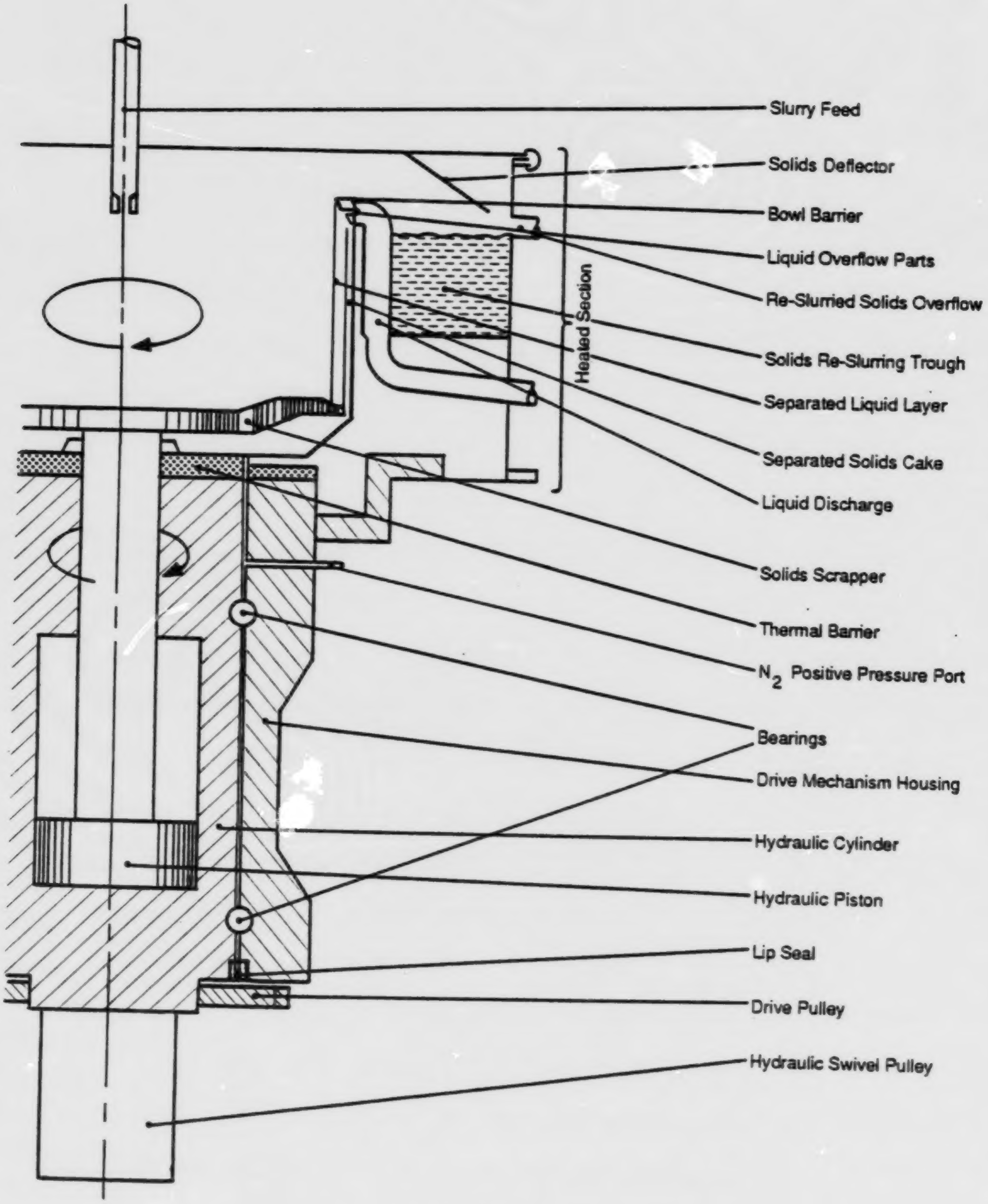

Slurry Feed
Solids Deflector
Bowl Barrier
Liquid Overflow Parts
Re-Slurried Solids Overflow
Solids Re-Slurring Trough
Separated Liquid Layer
Separated Solids Cake
Liquid Discharge
Heated Section
Solids Scrapper
Thermal Barrier
N_2 Positive Pressure Port
Bearings
Drive Mechanism Housing
Hydraulic Cylinder
Hydraulic Piston
Lip Seal
Drive Pulley
Hydraulic Swivel Pulley

FIGURE 4-1
SCHEMATIC OF CENTRAX SEPARATOR

layers increase in size until the liquid escapes from the bowl through the liquid overflow ports. From these ports liquid flows out of the liquid discharge channel and into the product receiving vessel 3.

After a certain time interval, the feed is terminated and the solids scraper moves upward at a pre-set speed while rotating at the same speed as the bowl. Both the residual liquid and solids are pushed upward by the scraper assembly: the liquid exits through the liquid ports, and the solids are pushed out of the bowl wall over the barrier at the top of the bowl when the scraper comes to the end of its stroke. The solids leave the spinning bowl tangentially and hit the solids ejector ring and are directed downwards into the solids re-slurrying trough which may either be empty (solids mode) or filled with extractant (slurry mode). This latter configuration would be used in a multi-staged extraction/separation operation.

Wear on the bowl wall is prevented by allowing a layer of solids cake one millimeter thick to build up on the wall. This was accomplished by designing the scraper with a deliberate clearance of one millimeter. Thus any relative motion is only between the scraper and the solids rather than with the wall. The bottom of the bowl wall is angled at 45°. This is designed to guide solids that may get under the scraper assembly back in front of the scraper.

4.2 Prototype Performance and Results

The CENTRAX prototype unit was commissioned in December 1984 and ten runs were made between that date and July 1985. Run times were generally between half an hour and one and one half hours. Thus the total data base generated in the seven months was eight hours of on-stream time.

Sandwell identified three main operational problem areas during execution of the test program:

- micro-flow controller and process flow meter problems
- temperature control problems in certain areas of the pilot plant
- bottom bearing problems in the CENTRAX separator unit

The last problem, the poor operation of the lower bearing, was a continuing problem throughout the test program. Several temporary measures were employed to ameliorate this, but Sandwell recognized that the section would have to be redesigned if an extensive development program was to be undertaken.

The test program was conducted using coal liquefaction materials obtained from the Wilsonville pilot plant. The feed material was an atmospheric flashed slurry product and the extractant was a solids-free hydrotreated bottoms liquid. The solids-containing material was a solid at room temperature with a softening point of about 80°C. At 120°C it is liquid, but during the phase change this material is "putty" like and difficult to handle. The extractant is liquid at room temperature with a low viscosity and is 100 percent oil (i.e., hexane soluble). The overall feed properties are given in Table 4-1.

The test program objectives were to determine the capabilities of the unit in effecting a separation of this coal derived material to produce a low solids overflow and a compacted solids underflow. Apparently the test program did not attempt to simulate multi-stage successive extractions but

TABLE 4.1
FEED AND EXTRACTANT COMPOSITION

COMPONENTS	FEED (wt %)	EXTRACTANT (wt %)
Oils	51.09	100
Asphaltenes	23.59	-
Pre-Asphaltenes	16.05	-
Solids	9.32	-
Unconverted Coal	4.45	-
Ash	4.87	-

concentrated on single stage processing of material.

The process variables studied were:

- feed flow rate (i.e., residence time)
- G force
- type of feed
- operating sequence of scraper cycle

During the test run, samples were taken of the separated liquid and the feed. These samples were routinely analyzed for ash, unconverted coal, preasphaltenes, asphaltenes, and oils. At the completion of a run additional samples were taken from vessels 3 and 4, the separated solids samples were also obtained from the CENTRAX trough and by scraping sections of the retained cake from the separator bowl wall. For some of the CENTRAX runs, simultaneous bench-scale runs were also performed with identical feeds to provide a comparison between the prototype and the bench-scale performance.

Sandwell reports test run data in "so-called detail" in Appendices 1 and 2 of the Phase III report (Sandwell, Phase III, 1985). Appendix 1 is essentially an operations log sheet that shows feed rates for various times during the run. Appendix 2 contains sporadic analyses of the separated liquid and separated solids at various times from the start of the run. Also for the majority of the runs, the overall analysis of the separated liquids from vessel 3 at the end of the run was documented.

Sandwell has performed an analysis of the results and have presented ash content of separated liquid as a function of feed rate and G force for the test runs 1 through 10. These data, although sparse, are represented graphically by Sandwell in Figure 4-2. The feed was a mixture of slurry product V110 material and T105 in the ratio of 1 to 0.3. The average ash content of the feed was 3.78 wt. percent. Data points, four in all, are represented by solid circles. Figure 4-2 shows clearly that in order to obtain low ash levels in the separated liquid, rates around 1 liter/minute are necessary. Higher flowrates (i.e., shorter residence times for separation) result, not surprisingly, in higher residual ash levels in the overflow. The effect of G between 2000 and 3000 appears to be minor compared to flowrate dependence. It should be noted, however, that ash is not the only solid present in this system. Unconverted coal (UC) is present and should also be removed from the separated liquid. Analysis of the data for the CENTRAX runs (Appendix 2) shows clearly that ash is preferentially removed relative to UC in this system. This observation is not surprising in view of the density differences between ash and unconverted coal particles. Figure 4-2 also shows an attempt to deash the separated liquid from run 9 (open squares). The apparent insensitivity of ash content to flowrate in this case is explained by Sandwell as being a function of solids particle size.

Another observation concerns the differences between analyses of reported liquid samples during the run and the analysis of the separated liquid in vessel 3 at the end of the run. Table 4-2 summarizes the CENTRAX run results and shows the end of run (EOR) UC and ash analyses of the separated liquid compared to the feed concentrations. Notice that the percentages of UC and ash in the separated liquid at EOR are several times

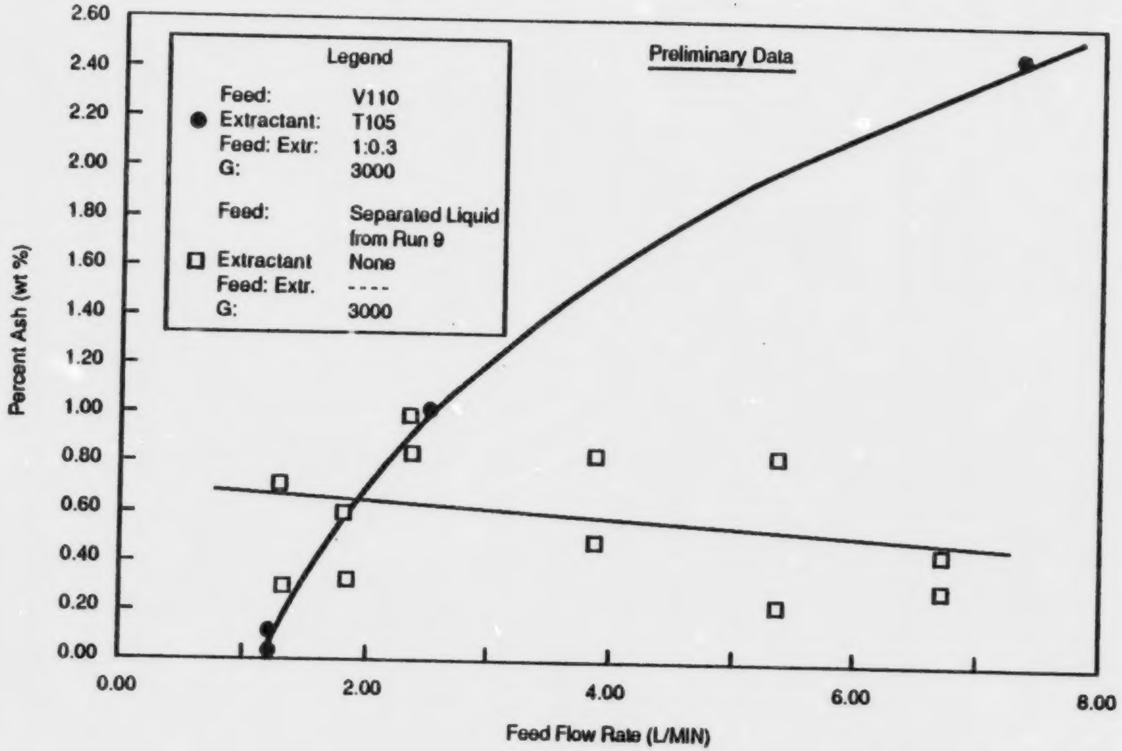


FIGURE 4-2
ASH CONTENT SEPARATED LIQUID
(TYPE OF FEED-FEED RATE)

PROPRIETARY INFORMATION

TABLE 4-2
SUMMARY OF CENTRAX RUN RESULTS

Run #	Time (Mins.)	Feed Rate (1/Min.)	G Force	EOR Sep Liquid		Feed	
				UC	Ash	UC	Ash
1	94	—	1180	0.69	0.62	2.47	3.03
2	63	0.9	1810	0.29	0.13	0.25	0.40
3	26	1.1	2710	0.87	0.83	2.91	2.89
4	43	0.9	2710	--	--	4.36	4.12
5	74	1.0	1910	2.23	1.01	2.78	1.85
6	39	1.0	2020	1.74	1.59	2.80	3.25
7	30	1.0	2020	3.41	1.25	2.46	2.50
8	29	0.9	2960	1.68	1.20	3.08	3.65
9	51	1-7	2960	2.03	0.95	2.99	3.78
10	30	1-7	2960	1.34	0.49	1.76	1.28

EOR = End of Run

UC = Unconverted Coal

that reported from sampling during the run.

A possible explanation for this is that the scraper pushes some solids through the liquid exit ports during its upstroke. Sandwell also suspected that this may occur and performed run 10 with the objective of determining if the operation of the scraper had an effect on the ash content of the separated liquid. By sampling the liquid overflow during different positions of the scraper during the cycle, Sandwell observed that slugs of solids occasionally came through with the liquid and that, upon inspection of the liquid discharge channel, solids were found within the channel. Sandwell suggested that during each upstroke of the scraper, a small quantity of solids were pushed through the liquid ports and this ended up in the separated liquids vessel. Sandwell concluded that redesign of that section of the unit be undertaken to minimize this effect of "solids cross-over" into the liquid product stream. No suggestions as to how this redesign would be accomplished were given in the report.

During CENTRAX prototype runs 7, 8, and 9 simultaneous laboratory centrifuge separations were conducted. The feed used was the same as that fed to the prototype unit. The laboratory test conditions were identical to those described in Section 3 of this report (i.e., separation temperature 180°C, spin time 5 minutes). Table 4-3 shows the comparative results of these experiments for run 7.

It is difficult to compare these two systems with respect to performance at a certain set of operating conditions since Sandwell has not provided specific information on the important parameters of the

TABLE 4-3
COMPARISON BETWEEN PROTOTYPE AND BENCH-SCALE UNITS - SEPARATED LIQUID CHARACTERISTICS RUN 7
(wt. percent)

PROTOTYPE

Time After Start of Run (Mins)	O	A	PA	UC	ASH	$\frac{O}{A + PA}$
0 (FEED)	62.39	20.31	11.93	2.81	2.57	1.94
3	69.66	16.86	11.89	1.19	0.40	2.42
6	66.24	17.44	15.15	1.17	0.17	2.03
8	70.99	16.35	11.33	1.17	0.16	2.56
12	66.20	17.99	13.96	1.59	0.28	2.10
15	65.24	20.54	12.94	1.13	0.14	1.95
Ave. Sep Liquid	67.66	17.84	13.05	1.22	.23	2.19

BENCH-SCALE

Sample 1	65.88	21.93	11.02	1.02	0.05	2.00
Sample 2	68.31	19.22	11.62	0.79	0.06	2.22
Sample 3	62.63	23.99	12.37	0.97	0.04	1.72

Note: All prototype samples are from liquid overflow pipe.

CENTRAX unit, i.e., the actual cycle time of the unit from when the feed is stopped, the scraper cycle completed and the feed restarted. One can obtain this information from run 10 (data given - Appendix 1, Phase III report) where the objective was to relate scraper operation to clarity of liquid overflow. In this run, cycle times varied from 2 minutes 30 seconds to 3 minutes 30 seconds; however, it is not clear what the cycle time needs to be to obtain acceptable performance during operations. This is a key parameter and must be a strong function of the feed viscosity and the density difference between the solids and liquid. In addition, it must be a function of the consistency of the solids cake and how this responds to the scraping action of the piston-scraper.

Referring to Table 4-3, it can be seen that levels of UC in the separated liquid are always between 4 and 8 times the level of ash for the prototype unit and about 20 times higher than ash levels for the laboratory centifuge. This observation was made earlier and relates to an inherent problem when using centrifugal devices for separating solids having such different specific gravities especially when the density difference is small between the liquid and one of the solids. The other observation from Table 4-3 concerns the relative ash levels in the separated liquids from the two systems. It appears that the laboratory centrifuge has a 4 times higher ash removal efficiency than the prototype. The reason for this is not obvious since the relative residence times in the two units are not clear.

The question as to whether A and PA material is selectively removed under centrifugation cannot be unequivocally answered, although from the ratio of 0/A&PA it appears that some selective rejection does occur. This is detrimental to product recovery. However, if these materials are rejected it may be advantageous from a process viewpoint since these materials are the most refractory.

Sandwell also investigated the characteristics of the separated solids from both the prototype and laboratory units. Table 4-4 shows the weight percent of components in samples obtained from various positions in the CENTRAX unit and from the residues in the laboratory centrifuge shield. The CENTRAX unit appears to compact the solids more effectively than the laboratory centrifuge as measured by the weight percent solids in the concentrate (i.e., 40-46 wt percent solids compared to 30-36 wt percent for the laboratory unit). This is probably because, in the CENTRAX unit, the upward action of the scraper imparts additional compression onto the solids cake. An additional reason could be that the liquid is decanted from the solid under 1 G in the laboratory tests, whereas in the CENTRAX unit the liquid and solids leave the unit under a 2000-3000 G force.

TABLE 4-4
SEPARATED SOLIDS COMPACTION -- PROTOTYPE COMPARED TO BENCH SCALE
(wt. percent)

PROTOTYPE

Sample Location	O	A	PA	UC	ASH	% Solids
Top of Bowl	32.24	6.85	20.33	12.90	27.68	40.58
	29.67	5.65	12.32	20.15	32.21	52.36
Middle of Bowl	34.52	17.42	3.53	21.06	23.47	44.53
Bottom of Bowl	33.59	6.86	15.00	19.94	24.61	44.55
Solids Trough	40.39	10.46	14.80	15.65	18.70	34.35
Ave.	34.08	9.45	13.20	17.94	25.33	43.27

BENCH-SCALE

Sample 1	40.54	11.09	11.42	17.14	19.81	36.95
Sample 2	44.23	11.36	15.34	13.64	15.43	29.07
Sample 3	40.06	15.23	9.21	16.14	19.36	35.50

5.0 MITRE ANALYSIS OF CENTRAX PERFORMANCE

5.1 Pilot Plant Performance

The most detailed results of the CENTRAX Pilot Plant presented by Sandwell are for run 7. MITRE has used these results to make an overall estimate of the performance of the CENTRAX unit. Run 7 was conducted at 2000g with an average flow rate of 0.93 liters/minute.

Tables 4-3 and 4-4 in the previous section show several measurements of overflow and separated solids composition for run 7. The average of the several measurements are presented in Table 5-1. Table 5-1 also shows the average characteristics of the overflow and separated solids as a percent of feed based on forced mass balances of ash and unconverted coal (UC). The results for forced ash and UC balances are very similar. These results indicate that approximately 60 percent of the UC and 92.5 percent of the ash present in the feed was recovered in the underflow during CENTRAX operations. The compacted underflow is 43.3 percent solids. These results are based on samples of the overflow during the run and compacted solids in the unit at the end of the run. These results make no account of the problem of solids crossover to overflow during scraper operation. Thus, the results are believed to be indicative of the level of performance which could be expected if the scraper problems discussed previously can be resolved in subsequent designs.

5.2 Multi Stage Performance

As a part of this study, MITRE has written a computer routine which uses iteration to estimate the flows through one, two, or three stages of

TABLE 5-1
OVERALL CENTRAX PERFORMANCE RUN 7 @ 2000g

	Oil	A	PA	UC	Ash	A+PA Oils
Measured Compositions						
Feed wt %	62.39	20.31	11.93	2.81	2.57	.517
Overflow wt %	67.67	17.84	13.05	1.22	0.23	.456
Separated Solids wt %	34.08	9.45	13.20	17.94	25.33	.66
Composition as Percent of Feed, Forced Ash Balance:						
Overflow	61.38	16.18	11.84	1.11	.21	
Separated Solids	3.18	.88	1.23	1.67	2.36	
Total	64.56	17.06	13.07	2.78	2.57	
Composition as Percent of Feed, Forced UC Balance						
Overflow	61.23	16.14	11.81	1.10	.21	
Separated Solids	3.29	.90	1.26	1.71	2.41	
Total	64.47	17.04	13.07	2.81	2.62	

centrifugation in the counter-current flow configuration proposed by Sandwell.

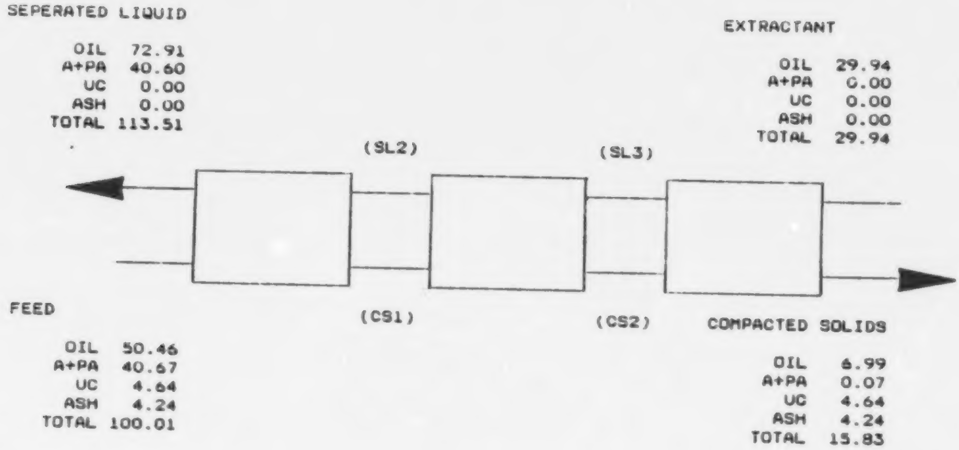
The required inputs to the model are the definition of the input streams of feed and extractant, an estimate of the percent solids in the compacted cake, and estimates of the percent of the ash and UC recovered in the compacted cake. The analysis assumes that the input materials to each stage are thoroughly mixed and that there is no centrifugal separation of soluble species. The outputs of each stage of centrifugation are thus assumed to have the same relative percentages of oils preasphaltenes (PA) and asphaltenes (A) as are present in the combined materials input to each stage.

The quantity of material in the underflow of a stage is determined by dividing the weight of solids recovered by the fractional solids content of the compacted cake.

The program uses the above relationship to compute flows through each stage. The first iteration is based on input estimates of the quantity of A and PA in the final underflow. This estimate is adjusted iteratively until the A and PA present in the underflow and overflow streams is equal to the A and PA introduced with the feed.

Figure 5-1 compares the results of the MITRE program with a Sandwell estimate of theoretical three stage separation presented in the Phase III report. The Sandwell analysis assumed 100 percent recovery of ash and UC, and 53.6 wt percent solids in the compacted cake. These values were input to the MITRE analysis. The results of the two analyses are very similar.

SANDWELL ESTIMATE


SEPERATED LIQUID
OIL 72.91
A+PA 40.60
UC 0.00
ASH 0.00
TOTAL 113.51
EXTRACTANT
OIL 29.94
A+PA 0.00
UC 0.00
ASH 0.00
TOTAL 29.94
(SL2)
(SL3)
FEED
OIL 50.46
A+PA 40.67
UC 4.64
ASH 4.24
TOTAL 100.01
(CS1)
(CS2)
COMPACTED SOLIDS
OIL 6.99
A+PA 0.07
UC 4.64
ASH 4.24
TOTAL 15.83

MITRE PROGRAM

```
CENTRAX DEASHER PERFORMANCE ANALYSIS
      COUNTERFLOW MODE
PERCENT UC REMOVED=              1.000
PERCENT ASH REMOVED =            1.000
PERCENT SOLIDS IN UNDERFLOW      0.561
```

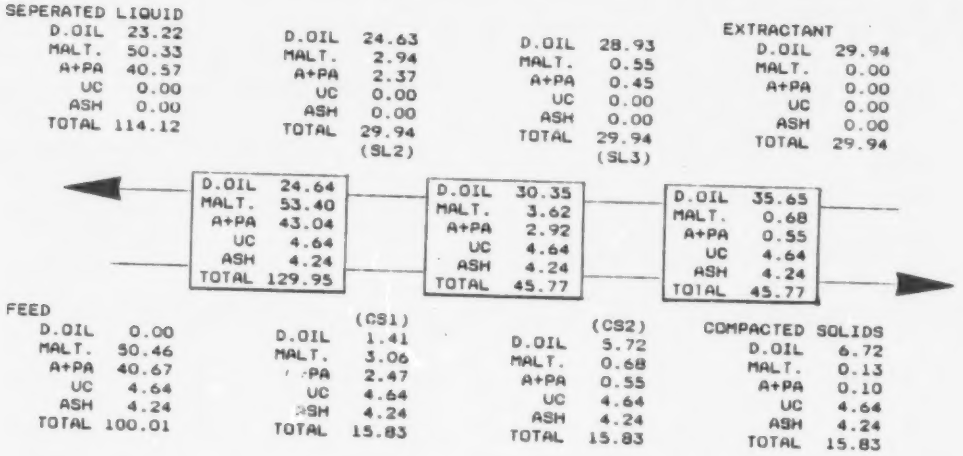

SEPERATED LIQUID
D.OIL 23.22
MALT. 50.33
A+PA 40.57
UC 0.00
ASH 0.00
TOTAL 114.12
D.OIL 24.63
MALT. 2.94
A+PA 2.37
UC 0.00
ASH 0.00
TOTAL 29.94
(SL2)
D.OIL 28.93
MALT. 0.55
A+PA 0.45
UC 0.00
ASH 0.00
TOTAL 29.94
(SL3)
EXTRACTANT
D.OIL 29.94
MALT. 0.00
A+PA 0.00
UC 0.00
ASH 0.00
TOTAL 29.94
D.OIL 24.64
MALT. 53.40
A+PA 43.04
UC 4.64
ASH 4.24
TOTAL 129.95
D.OIL 30.35
MALT. 3.62
A+PA 2.92
UC 4.64
ASH 4.24
TOTAL 45.77
D.OIL 35.65
MALT. 0.68
A+PA 0.55
UC 4.64
ASH 4.24
TOTAL 45.77
FEED
D.OIL 0.00
MALT. 50.46
A+PA 40.67
UC 4.64
ASH 4.24
TOTAL 100.01
(CS1)
D.OIL 1.41
MALT. 3.06
A+PA 2.47
UC 4.64
ASH 4.24
TOTAL 15.83
(CS2)
D.OIL 5.72
MALT. 0.68
A+PA 0.55
UC 4.64
ASH 4.24
TOTAL 15.83
COMPACTED SOLIDS
D.OIL 6.72
MALT. 0.13
A+PA 0.10
UC 4.64
ASH 4.24
TOTAL 15.83

FIGURE 5-1
THEORETICAL ANALYSIS OF THREE STAGE COUNTERFLOW CENTRAX SEPARATION AND EXTRACTION

The total A and PA in the cake in the MITRE analysis is .10 vs .07 in the Sandwell analysis. Note that the MITRE analysis differentiates between distillable oils and maltenes (nondistillable oils).

The MITRE results show the quantity of flow in each stage of centrifugation. Note that only stage 1 handles a flow equivalent to the total of feed and extractant. The flow in the subsequent stages is equal to the quantity of extractant plus the cascading underflow. The quantity of cascading underflow is equal to the quantity of solids present divided by the percent solids in the compacted cake.

Sandwell's laboratory simulation of multi-stage centrifugation was discussed in Section 4. Figure 5-2 compares the results of this MITRE program with these results. The percent solids in the compacted cake, and the percent of ash and UC recovered in the tests were input to the MITRE program. In the program, the percent ash and UC recovered was varied iteratively until the quantities in the final underflow agreed with the test results. The analysis shown assumes that equal percentages of ash and UC are lost to the overflow in each stage of centrifugation.

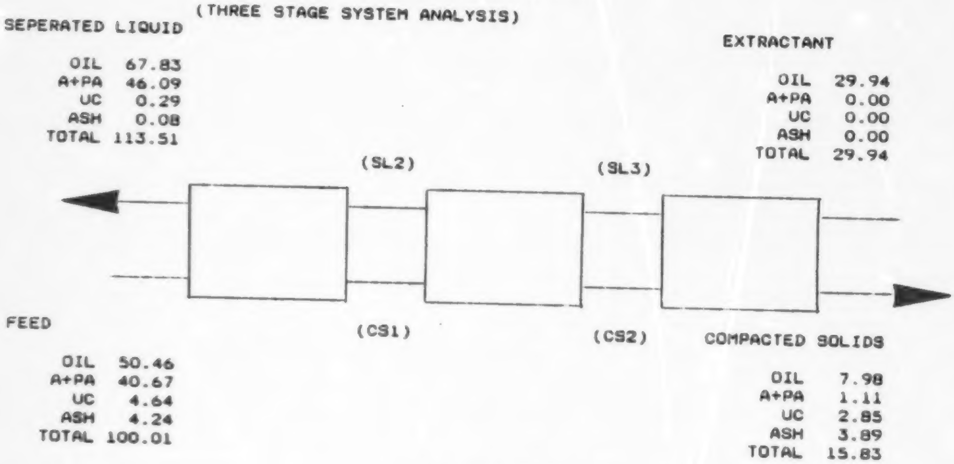
In practice, the loss of solids to the overflow would occur principally in the first stage, since fluid viscosity is lower in subsequent stages, and solids too small to be recovered in the first stage have already been lost to the system.

The forced material balance of the MITRE analysis demands that solids not present in the underflow be accounted for in the overflow. A and PA must also balance. The overflow estimated by the MITRE program is, thus,

SANDWELL TEST RESULTS

```
CENTRAX DEASHER PERFORMANCE ANALYSIS
      COUNTERFLOW MODE
PERCENT UC REMOVED=                 0.714
PERCENT ASH REMOVED =               0.923
PERCENT SOLIDS IN UNDERFLOW         0.426

(THREE STAGE SYSTEM ANALYSIS)
```


SEPERATED LIQUID
OIL 67.83
A+PA 46.09
UC 0.29
ASH 0.08
TOTAL 113.51
EXTRACTANT
OIL 29.94
A+PA 0.00
UC 0.00
ASH 0.00
TOTAL 29.94
(SL2)
(SL3)
FEED
OIL 50.46
A+PA 40.67
UC 4.64
ASH 4.24
TOTAL 100.01
(CS1)
(CS2)
COMPACTED SOLIDS
OIL 7.98
A+PA 1.11
UC 2.85
ASH 3.89
TOTAL 15.83

MITRE PROGRAM

```
CENTRAX DEASHER PERFORMANCE ANALYSIS
      COUNTERFLOW MODE
PERCENT UC REMOVED=                 0.714
PERCENT ASH REMOVED =               0.923
PERCENT SOLIDS IN UNDERFLOW         0.426

(THREE STAGE SYSTEM ANALYSIS)
```

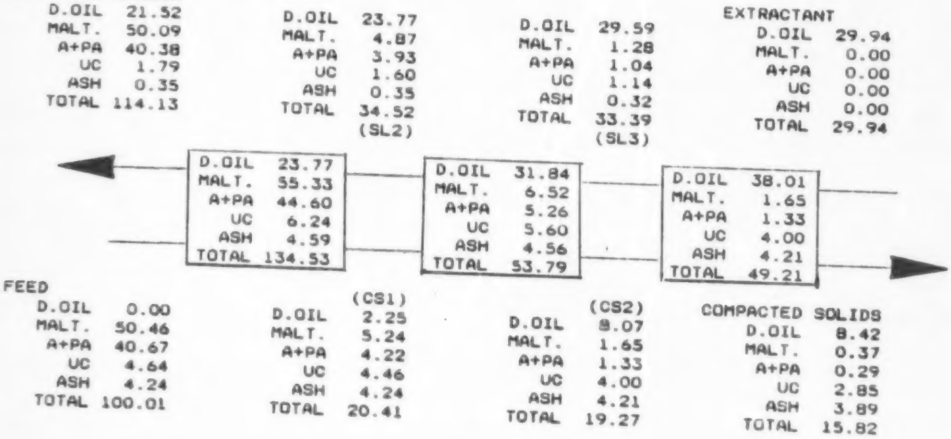

SEPERATED LIQUID
D.OIL 21.52
MALT. 50.09
A+PA 40.38
UC 1.79
ASH 0.35
TOTAL 114.13
D.OIL 23.77
MALT. 4.87
A+PA 3.93
UC 1.60
ASH 0.35
TOTAL 34.52
(SL2)
D.OIL 29.59
MALT. 1.28
A+PA 1.04
UC 1.14
ASH 0.32
TOTAL 33.39
(SL3)
EXTRACTANT
D.OIL 29.94
MALT. 0.00
A+PA 0.00
UC 0.00
ASH 0.00
TOTAL 29.94
D.OIL 23.77
MALT. 55.33
A+PA 44.60
UC 6.24
ASH 4.59
TOTAL 134.53
D.OIL 31.84
MALT. 6.52
A+PA 5.26
UC 5.60
ASH 4.56
TOTAL 53.79
D.OIL 38.01
MALT. 1.65
A+PA 1.33
UC 4.00
ASH 4.21
TOTAL 49.21
FEED
D.OIL 0.00
MALT. 50.46
A+PA 40.67
UC 4.64
ASH 4.24
TOTAL 100.01
(CS1)
D.OIL 2.25
MALT. 5.24
A+PA 4.22
UC 4.46
ASH 4.24
TOTAL 20.41
(CS2)
D.OIL 8.07
MALT. 1.65
A+PA 1.33
UC 4.00
ASH 4.21
TOTAL 19.27
COMPACTED SOLIDS
D.OIL 8.42
MALT. 0.37
A+PA 0.29
UC 2.85
ASH 3.89
TOTAL 15.82

FIGURE 5-2
SANDWELL SIMULATION AND MITRE ANALYSIS OF THREE STAGE COIUNTERFLOW CENTRAX SEPARATION

much different from the overflow produced in the tests.

The differences in underflow are more subtle but perhaps more significant. Ash, UC, and total liquids values agree with the tests since test values were used as input. However, under the assumption of perfect mixing and no preferential centrifugation of A and P materials, the A and PA present in the underflow is seen to be .29 versus the 1.11 achieved in the tests. This difference may indicate that there is a significant selective rejection of heavier A and PA materials in the underflow in the test results. We noted earlier that some selective precipitation of the A and PA would be expected to the extent that these species are not soluble in the extract. A relatively small concentration of A and PA material in each stage can result in a significant concentration in the final underflow. Asphaltenes and preasphaltenes are not specific molecular entities but span a wide range of molecular weights. Heavier species, as well as those intimately bound to solid particles may resist homogeneous mixing. If such molecules account for about 20 percent of the A and PA in the feed, then final concentrations of A and PA three times the levels predicted by perfect mixing assumptions could occur in the underflow.

We have noted earlier that the percent rejection of ash in the underflow is higher than the rejection of unconverted coal. The impact of this observation is problematical. If the unconverted coal in the overflow is dissolved or converted when it is recycled to the liquefaction system, then its retention in the overflow is advantageous. If it is not converted in subsequent passes, then the UC in the recycle stream will build up to the point that the percent reject equals that produced by the liquefaction system on a steady state basis. These considerations are addressed subsequently when the MITRE counterflow model is used to estimate CENTRAX performance in the 10,000 b/day commercial plant considered in the Sandwell Phase III report.

5.3 Scaling Considerations

5.3.1 Sigma Theory Scaling

The report written by Professors Wade and Judd of McMaster University briefly discusses CENTRAX scale up information (McMaster, Wade and Judd, 1986). On the advice of the prototype designers (DSMA International) and with consultation with Sandwell, the scaled up version of the CENTRAX unit would consist of an optimum bowl diameter of 950 mm based on stress analysis calculations assuming a particle acceleration requirement of 3000 g, a cake thickness of 5mm and operating at a rotational speed of 2440 rpm. No estimate of scaled-up bowl length was given in the report. Using the same L/D ratio as the CENTRAX prototype unit, the commercial scale length would be approximately 570 mm.

The theory relating the throughput of a centrifuge for a given performance to the physical properties of the fluid/solid system and geometrical parameters of the centrifuge is thoroughly substantiated in practice (Bradley, 1965 and Lavanchy, 1979). This particular approach is known as the "Sigma Theory."

In the general case, it can be shown that:

$$Q = 2Vg\,\Sigma$$

where $\Sigma = \frac{Vw^2r}{gS}$ and $Vg = \frac{D^2g\Delta\rho}{\eta 18}$

In this context the following nomenclature apply:

Q = Volumetric flow of process liquid through centrifuge in cc/second

Vg = Terminal velocity of solid particle settling under gravity through liquid in cm/sec.

Σ = Sigma value of centrifuge in sq. cm.

V = Volume of liquid in centrifuge bowl

W = Angular speed of rotation in radians per second

r = Radius of bowl wall

S = effective setting distance

$\Delta\rho$ = difference in density between particles and medium

D = diameter of particle half of which is removed in passage through centrifuge (known as the 'cut point')

η = viscosity of liquid phase in poises

r_e = radius of the free surface of the liquid in the cylindrical centrifuge

For separations in an imperforated bowl certrifuge of tubular design it can further be shown that:

$Q = 2\ Vg\Sigma_T$ where

$$\Sigma T = 2\pi \frac{Lw^2}{g} (3/4\ r^2 + 1/4\ r_e^2)$$

Using these equations, the Σ_T value of the CENTRAX prototype unit can be estimated. The assumed value of r_e is one centimeter less than r, since this is the height of the flange preventing overflow in the CENTRAX prototype. Thus:

$$\Sigma_T = 2\pi \; 15 \; \frac{(2\pi \; .76)^2}{981} \; (0.75\,(12.5)^2 + 0.25\,(11.5)^2)$$

$$\Sigma_T = \underline{3{,}293{,}000} \quad @ \text{ RPM} = 4600 \; (\,3000g)$$

and $$\Sigma_T = \underline{2{,}263{,}000} \quad @ \text{ RPM} = 3800 \; (\,2000g)$$

Now calculating Σ_T for the scaled up commercial CENTRAX unit we have:

$$\Sigma_T = 2\pi \times 57 \; \frac{(2\pi \times 40.8)^2}{981} \; (0.75\,(47.5)^2 + 0.25\,(46.5)^2)$$

$$\Sigma_T = \underline{53{,}583{,}000} \quad @ \text{ RPM} = 2440, \; (3000g)$$

Thus: $$\frac{\Sigma_T \text{ Commercial}}{\Sigma_T \text{ Prototype}} = 16.27 \quad @ \; 3000g$$

and $$\frac{\Sigma_T \text{ Commercial}}{\Sigma_T \text{ Prototype}} = 23.67 \quad @ \; 2000g$$

The factors that define Vg are determined by feed properties and separation performance. Thus for similar feeds and performance, the flow rate of full scale units at 3000g may be about 16 times the flow rate derived in the prototype at 3000g, or about 24 times the prototype flow rate at 2000 gs.

The most detailed results presented in the Phase III report are for run 7, which was operated at 2000g and a flowrate of .93 liters/minutes (15.5 cc/sec). For this run, the estimated Q/ is $15.5/2.26 \times 10^{6} = .686 \times 10^{-5}$, which is relatively low when compared to typical Q/ values for tubular centrifuges.

From the equation $Vg = \dfrac{g\Delta\rho \quad D^{2}}{18\eta}$

it is possible to estimate the theoretical "cut point" for the prototype CENTRAX unit. The cut point is the particle diameter half of which is removed by the centrifuge. The following feed properties are assumed.

$\Delta\rho$ Ash = 1.5 g/cc

$\Delta\rho$ U.C. = 0.3 g/cc

η of fluid = 10 cp

For the UC particles:

$$0.343 \times 10^{-5} = \frac{.3 \times 981 \times D^{2}}{1.8}$$

$D = 1.427 \times 10^{-4}$cm = 1.4 microns

For ash particles:

$$0.343 \times 10^{-5} = \frac{1.5 \times 981 \times D^{2}}{1.8}$$

$D = 0.6 \times 10^{-4}$cm = 0.6 microns

5.3.2 Solids Removal Considerations

If the commercial unit is sized as described above, the bowl volume would be 404 liters. The bowl volume of the prototype unit is 7.4 liters. A commercial unit would have a surface for solids deposition of 18.31 ft^2. If the allowable build up of cake is 5 mm in thickness, then each commercial unit has the capacity to remove 0.3004 ft^3 of compacted solids for each separator cycle.

Data given in the Phase III report, Appendix 1, summarize the experimental operating conditions of the prototype unit for runs 1 through 10. In the final reported run 10, precise run times are given for feed startup, feed stoppage, and scraper rise time. Figure 5-3 shows the cycle sequence for a portion of this run. Other periods of the run show cycle times for feed cessation to restart ranging from about 2 minutes 30 seconds to 3 minutes 30 seconds. Thus, an average cycle time would be in the order of 3 minutes and one commercial unit could remove 0.3004 ft^3 of compacted solids every 3 minutes, or approximately 0.1 ft^3/minute.

5.4 Performance of Full Scale CENTRAX Units

5.4.1 Estimated Number of Full Scale Units Required

In the Phase III report, Sandwell described the application of full scale CENTRAX separators to a conceptual 10,000 bbl/day liquefaction facility. The deasher feed rate is reported to be 5064

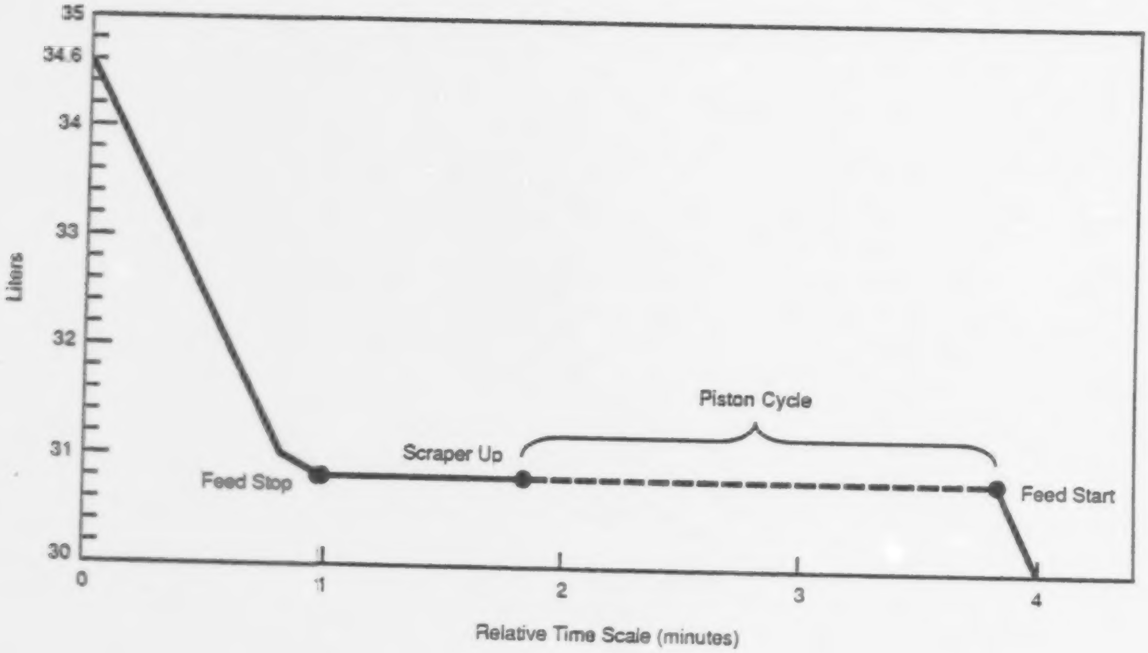


FIGURE 5-3
CENTRAX PROTOTYPE CYCLE TIMES FOR RUN #10

tons/day. Characteristics of the deasher feed material are given in Table 5-2.

The feed of 5064 tons/day of 1.069 SG materials represents a volumetric feed of 3000 liters per minute. Based on Sigma scaling, using run 7 (.93 liters/minute at 2000g) as the basis, the full scale units would process .93 x 23.67 = 22 liters/minute. Thus, 3000/22 = 136 units would be required for the first stage if flow were continuous. Subsequent stages process a flow that is approximately equivalent to the sum of compacted solids and extractant. At 40 percent compaction, the flow to subsequent stages is approximately 2889 t/d or approximately 1517 liters per minute. For continuous flow, the number of centrifuge in subsequent stages would be 1517/22 = 69. A three-stage CENTRAX system would, thus, require about 280 CENTRAX units operating continuously.

The CENTRAX units cannot operate continuously, however, because of the requirement to interrupt the flow during the action of the scraper. We noted earlier that each cycle of the scraper is capable of removing .3004 ft^3 of solids. As noted in Table 5-2, the commercial plant produces 1443 T/D of compacted solids having a specific gravity of 1.43. The solids volume is thus 21.8 ft^3/minute. If the CENTRAX unit has a cycle time of 3 minutes as illustrated in Figure 5-3, then the average solids removal rate is .1 ft^3/minute. Thus, 218 units would be required to remove the solids generated in the 10,000 bbl/d commercial plant. The same number of units would be required in each stage since the solids are compacted and removed in each stage. A total of 654 units would,

TABLE 5-2
ESTIMATED PROPERTIES OF DEASHER FEED
10,000 bbl/Day COMMERCIAL FACILITY

	T/D	Specific Gravity	Viscosity CP
ASH (.205 MAF Coal)	415	2.5	
UC (0.08 MAF Coal)	162	1.3	
SRC from fresh Coal (.409MAF)	989	1.0	10-20
SRC from Recycle (approx = MAF)	2051	1.0	10-20
Subtotal	3617		
Extracted at 1:0.4	1447	.94	1-2
Total (Average)	5069	(1.064)	(7.25-14.5)
Liquid Only (feed & extractant)		.98	7.25-14.5
Compacted Solids 40% Compacted	1443	1.43	
Compacted Solids & Extractant	2889	1.20	2-4

therefore, be required. Cycle times must be reduced to less than 3 minutes, or solids removal consideration will be much more limiting than centrifugation times in the determination of the number of units required.

Further insights into the relationship between the two scaling limitations can be obtained by considering the time of flow in each cycle that is compatible with the limiting solids build up of 5 mm.

We noted earlier that the 5 mm layer of compacted solids represented a volume of 0.3004 ft^3 in the full scale unit. At a compaction of 40 percent solid by weight, and a compacted solids specific gravity of 1.43, the weight of the solid in the compacted cake is 10.72 lbs. Since the flow to the centrifuge is 11.4 percent solids by weight, the total flow required to deposit 10.72 lbs of solid is 94.05 lbs, or 38.17 liters. If the flow to the unit is 22 liters/minute, as indicated by Sigma scaling, then flow time per cycle will be 1.77 minutes for the first stage. In subsequent stages, the flow is 2889 T/D of 20 percent solids material having a S.G of 1.2. The flow required to deposit 10.72 lbs of solids is thus 20.54 liters. Flow time @ 22 liters/minute would be 0.95 minutes.

The total cycle time for each unit will be the sum of flow time plus scraper cycle time. Figure 5-4 shows the total number of units required for the 10,000 b/d commercial plant as a function of scraper cycle time. Flow times used are 1.77 minutes for the first stage and 0.95 minutes for subsequent stages. The number of units varies from about 300 to about 500 as cycle time varies from 0 to 60 seconds. This relationship illustrates the significance of keeping the scraper cycle time to a minimum.

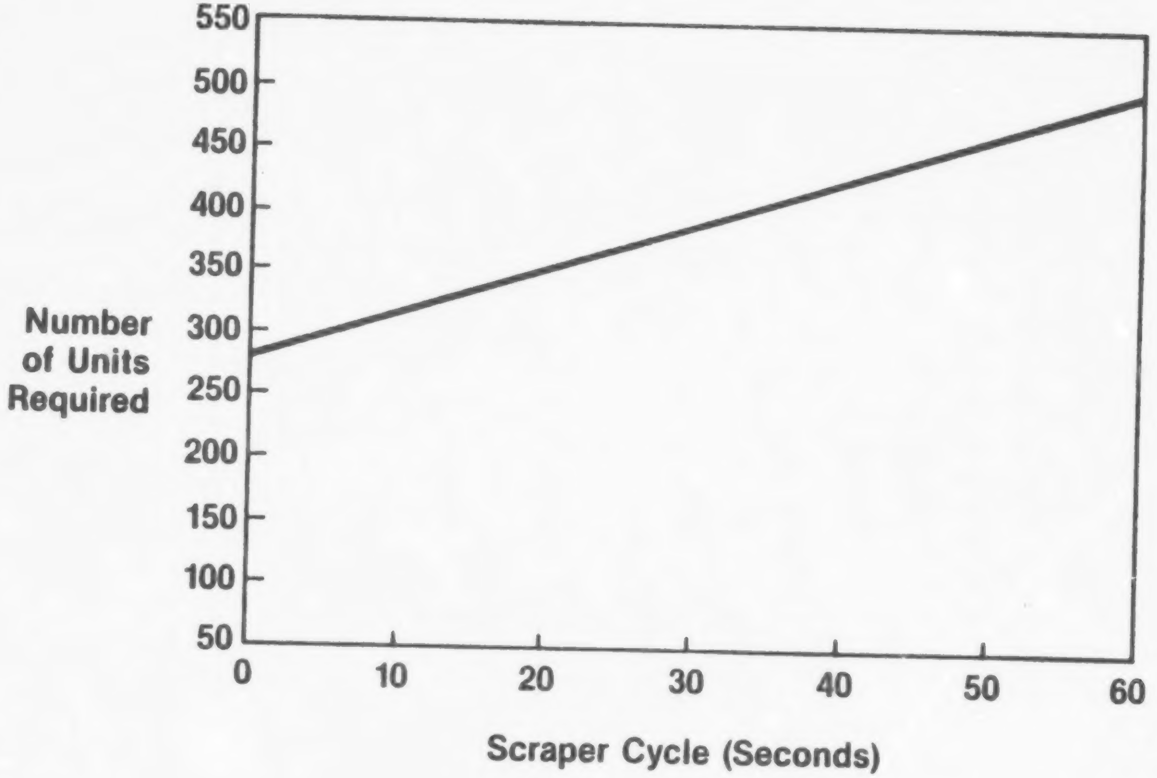


FIGURE 5-4
FULL SCALE UNITS REQUIRED VS. SCRAPER CYCLE TIME

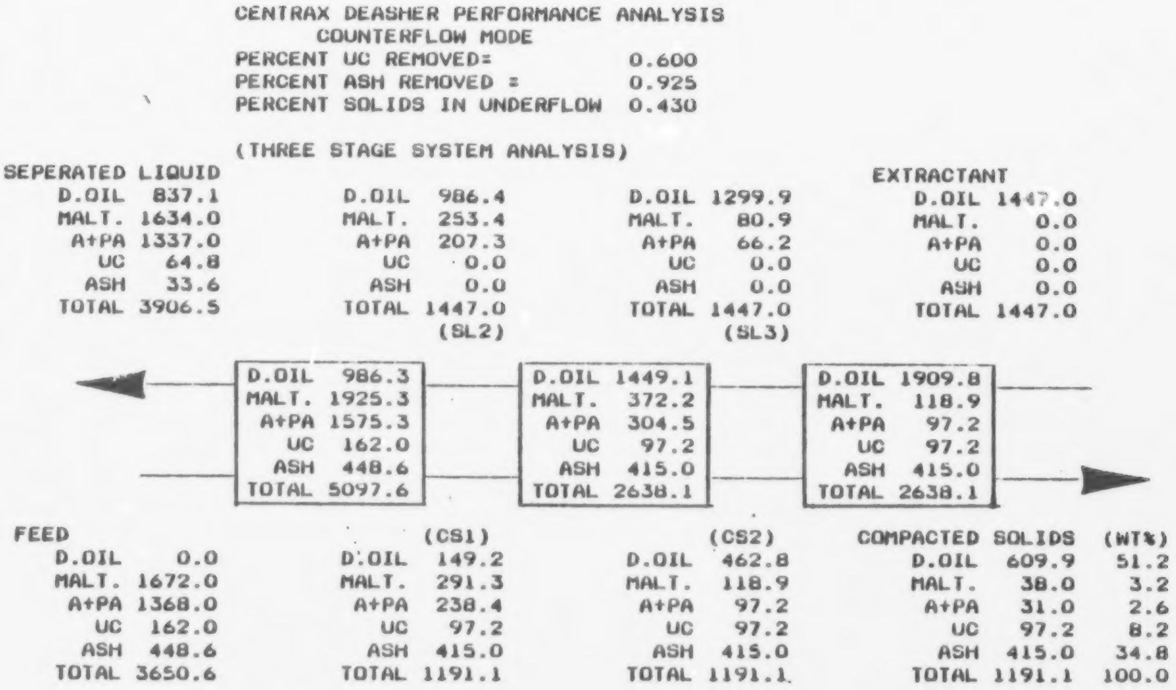
The Sandwell Reports do not present a specific projection of the scraper cycle time for the full scale unit. In earlier conversations, times in the range of 15 to 20 seconds were mentioned. At a scraper cycle time of 20 seconds, the 10,000 bbl/day commercial plant would require 350 CENTRAX units. This estimate appears to be consistent with the Sandwell estimate of $46 million for the cost of CENTRAX deashing in the plant analyzed. At $100,000 for each unit, 350 units will cost $35 million to which feed preparation and drying equipment costs must be added.

5.4.2 Estimated Performance

The computer program described in Section 5-2 has been used to estimate the performance of full scale CENTRAX units operating in the three stage counter-current mode. The feed and extractant are defined in Table 5-2.

Figure 5-5 shows the results of a baseline run of the three stage counterflow computer model. Inputs are derived form single stage results achieved in run 7 of the CENTRAX Prototype unit. Specifically, it is assumed that the compacted cake is 43 percent solids, and that 60 percent of UC and 92.5 percent of ash are recovered in the final underflow.

The units of output are tons/day except for the final compacted cake for which the wt. percent composition is also given. The loss of solids to the overflow is assumed to occur in the first stage only. The concentration of ash that results from the assumed 92.5 percent removal is accounted for by increasing input ash by 1/0.925. No concentration of UC is assumed. Results from this run represent an idealized case where UC in



```
          CENTRAX DEASHER PERFORMANCE ANALYSIS
               COUNTERFLOW MODE
          PERCENT UC REMOVED=            0.600
          PERCENT ASH REMOVED =          0.925
          PERCENT SOLIDS IN UNDERFLOW    0.430

          (THREE STAGE SYSTEM ANALYSIS)
SEPERATED LIQUID                                                EXTRACTANT
    D.OIL  837.1         D.OIL  986.4        D.OIL 1299.9          D.OIL 1447.0
    MALT. 1634.0         MALT.  253.4        MALT.   80.9          MALT.    0.0
     A+PA 1337.0          A+PA  207.3         A+PA   66.2           A+PA    0.0
       UC   64.8            UC    0.0           UC    0.0             UC    0.0
      ASH   33.6           ASH    0.0          ASH    0.0            ASH    0.0
    TOTAL 3906.5         TOTAL 1447.0        TOTAL 1447.0          TOTAL 1447.0
                                (SL2)                (SL3)

 <-------- D.OIL   986.3 ------ D.OIL 1449.1 ------ D.OIL 1909.8 --------
           MALT. 1925.3        MALT.   372.2       MALT.   118.9
            A+PA 1575.3         A+PA   304.5        A+PA    97.2
              UC  162.0           UC    97.2          UC    97.2
     ------  ASH  448.6 ------   ASH   415.0 ------  ASH   415.0 -------->
           TOTAL 5097.6        TOTAL  2638.1       TOTAL  2638.1

FEED                           (CS1)                (CS2)       COMPACTED SOLIDS  (WT%)
    D.OIL    0.0         D.OIL  149.2        D.OIL  462.8          D.OIL  609.9   51.2
    MALT. 1672.0         MALT.  291.3        MALT.  118.9          MALT.   38.0    3.2
     A+PA 1368.0          A+PA  238.4         A+PA   97.2           A+PA   31.0    2.6
       UC  162.0            UC   97.2           UC   97.2             UC   97.2    8.2
      ASH  448.6           ASH  415.0          ASH  415.0            ASH  415.0   34.8
    TOTAL 3650.6         TOTAL 1191.1        TOTAL 1191.1          TOTAL 1191.1  100.0
```

FIGURE 5-5
CENTRAX PERFORMANCE 10,000 B/Day PLANT:
CASE 1: UC IN OVERFLOW CONVERTS WHEN RECYCLED

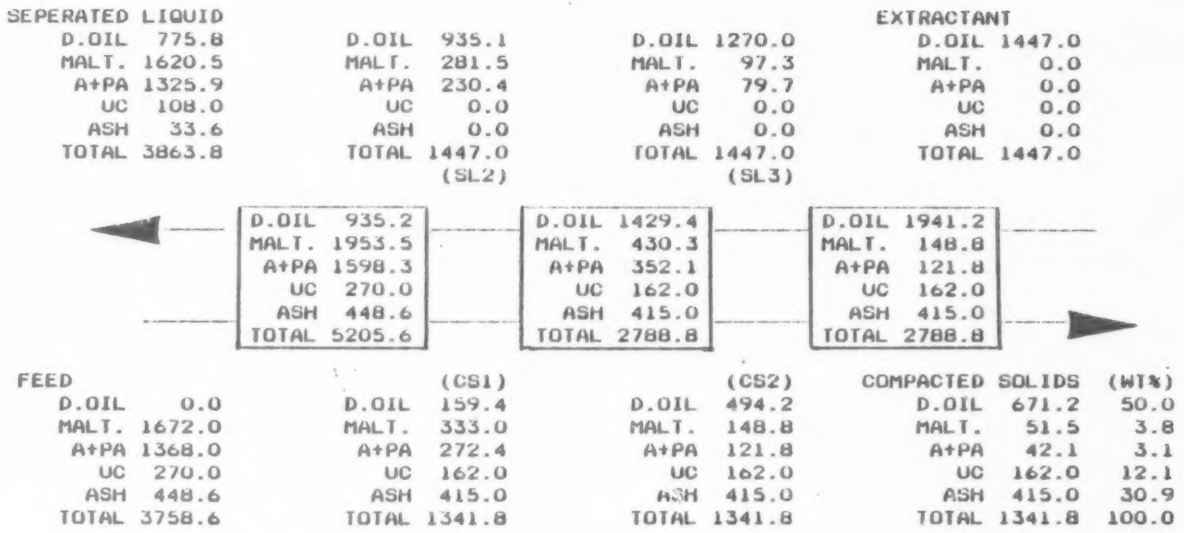
PROPRIETARY INFORMATION

the overflow is converted during recycle to the liquefaction unit. Under these ideal assumptions, 95.2 wt. percent solids in reject could be obtained if all oils and maltenes could be recovered in the extractant recovery process. This level of performance is almost identical to the 96 percent assumed by Sandwell in their economic scoping study.

Figure 5-6 shows the results of the MITRE CENTRAX model if the UC does not convert and builds up until steady state accumulation is removed. Under this assumption, a dried compacted cake containing 93.2 percent solids could be achieved if all material except A and PA could be recovered in the extractant recovery step. In practice we do not think it practical to remove maltenes during the procedure. Drying conditions severe enough to remove these non-distillate oils would be equivalent to coking and would lead to losses due to coke formation. Furthermore, some of the recovered material would most likely be refractory to additional liquefaction.

If the extractant recovery system could remove all of the distillable oils, then the dried compacted cake would contain 86.2 percent solids. This result is still optimistic since it makes no provision for possible concentration of A and PA during centrifugation. Additionally, some retention of distillate oils during the drying process can be expected. For example, in the CSD process, the ash concentrate still contains some solvent after the spray drying process and this solvent is considerably lower boiling then the CENTRAX extractant.

```
CENTRAX DEASHER PERFORMANCE ANALYSIS
      COUNTERFLOW MODE
PERCENT UC REMOVED=            0.600
PERCENT ASH REMOVED =          0.925
PERCENT SOLIDS IN UNDERFLOW    0.430
```



```
SEPERATED LIQUID                                                      EXTRACTANT
   D.OIL  775.8        D.OIL  935.1         D.OIL 1270.0          D.OIL 1447.0
   MALT. 1620.5        MALT.  281.5         MALT.   97.3          MALT.    0.0
    A+PA 1325.9         A+PA  230.4          A+PA   79.7           A+PA    0.0
      UC  108.0           UC    0.0            UC    0.0             UC    0.0
     ASH   33.6          ASH    0.0           ASH    0.0            ASH    0.0
   TOTAL 3863.8        TOTAL 1447.0         TOTAL 1447.0          TOTAL 1447.0
                             (SL2)                (SL3)

 <----  D.OIL   935.2      D.OIL 1429.4      D.OIL 1941.2
        MALT. 1953.5      MALT.  430.3      MALT.  148.8
         A+PA 1598.3       A+PA  352.1       A+PA  121.8
           UC  270.0         UC  162.0         UC  162.0
          ASH  448.6        ASH  415.0        ASH  415.0
        TOTAL 5205.6      TOTAL 2788.8      TOTAL 2788.8  ---->

FEED                             (CS1)                (CS2)   COMPACTED SOLIDS  (WT%)
   D.OIL    0.0        D.OIL  159.4         D.OIL  494.2        D.OIL  671.2   50.0
   MALT. 1672.0        MALT.  333.0         MALT.  148.8        MALT.   51.5    3.8
    A+PA 1368.0         A+PA  272.4          A+PA  121.8         A+PA   42.1    3.1
      UC  270.0           UC  162.0            UC  162.0           UC  162.0   12.1
     ASH  448.6          ASH  415.0           ASH  415.0          ASH  415.0   30.9
   TOTAL 3758.6        TOTAL 1341.8         TOTAL 1341.8        TOTAL 1341.8  100.0
```

FIGURE 5-6
CENTRAX PERFORMANCE IN 10,000 b/d PILOT PLANT
CASE 2: UC IN OVERFIOW CONCENTRATES

PROPRIETARY INFORMATION

Table 5-3 shows results for what are believed to be more realistic conditions. A and PA are increased to levels twice as high as are predicted using assumptions of perfect mixing as is assumed in the MITRE CENTRAX model. Also, 5 wt. percent of the distillable oil is assumed to remain in the dried cake. Under these assumptions the dried cake contains 77.6 percent solids. The soluble organic reject under these conditions is 22.4 percent compared to the 4 percent assumed by Sandwell in their economic analysis. However, this realistic level of performance is still excellent relative to CSD or other alternative deashing systems. Furthermore, the economic advantages, in terms of cost per barrel of liquids, of improved deashing beyond 75 to 80 percent solids accrue very slowly (See Figure 1-2).

Figure 5-7 graphically illustrates the composition of the dried compacted cake for the three assumptions in the CENTRAX model described above. The economic implications of using CENTRAX deashing at each of these three performance levels has been evaluated in Section 6 of this report.

5.4.3 Sensitivity of CENTRAX Performance to Cake Compaction

The analysis presented in the section above assumed a centrifugal compaction to 43 percent solids in the cake. This is the level of compaction achieved in both the bench-scale and prototype units. Figure 5-8 shows the effect of different compaction levels of the compacted cake on the solids content in the dried cake. These results are from the three-stage counter-current MITRE CENTRAX simulation model.

TABLE 5-3
COMPACTED AND DRIED CAKE COMPOSITION FOR REALISTIC CENTRIFUGATION AND DRYING ASSUMPTIONS

	Compacted Solids		Dried Cake	
	T/Day	Wt %	T/D	Wt%
D. Oils	629.1	46.2	31.5	4.2
Maltenes	51.5	3.8	51.5	6.9
A and PA	84.2	6.2	84.2	11.3
UC	162	12.1	162	21.8
Ash	415	30.9	415	55.8
Total	1341.8	100.0	744.2	100

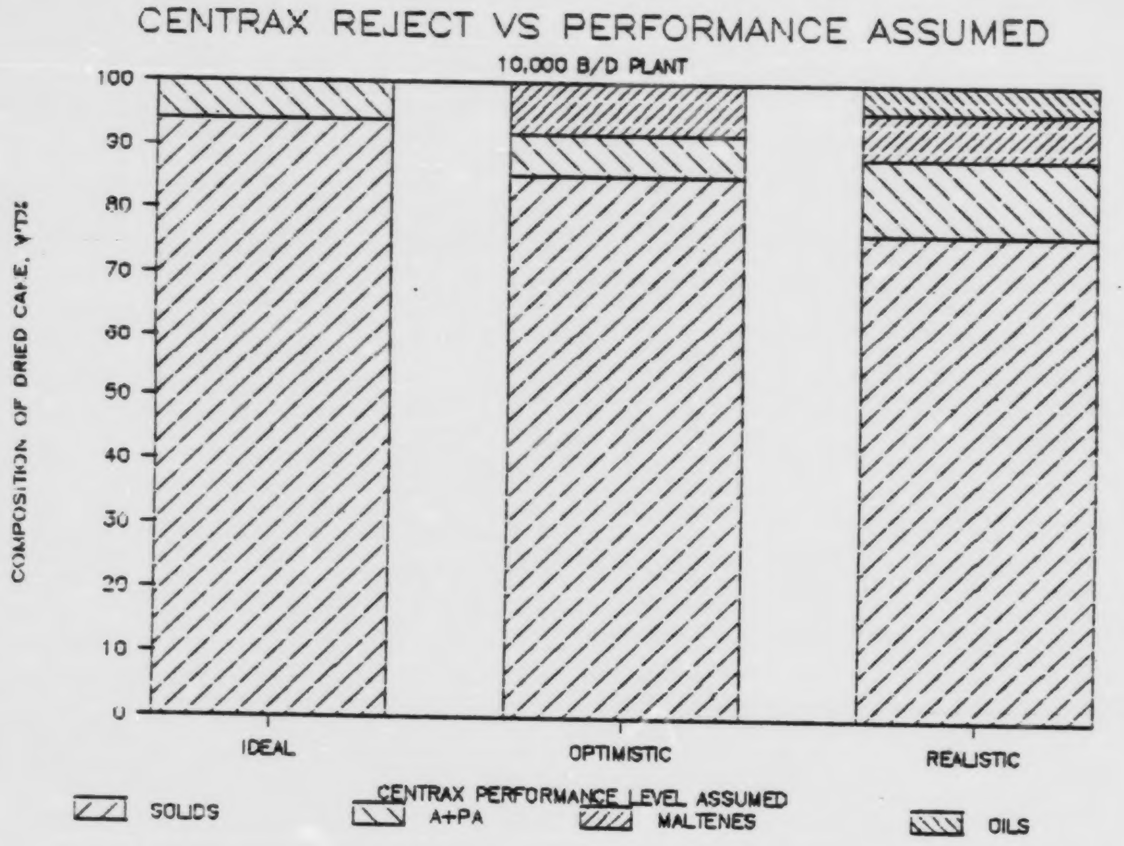


FIGURE 5-7
CENTRAX REJECT VS PERFORMANCE ASSUMED

PROPRIETARY INFORMATION

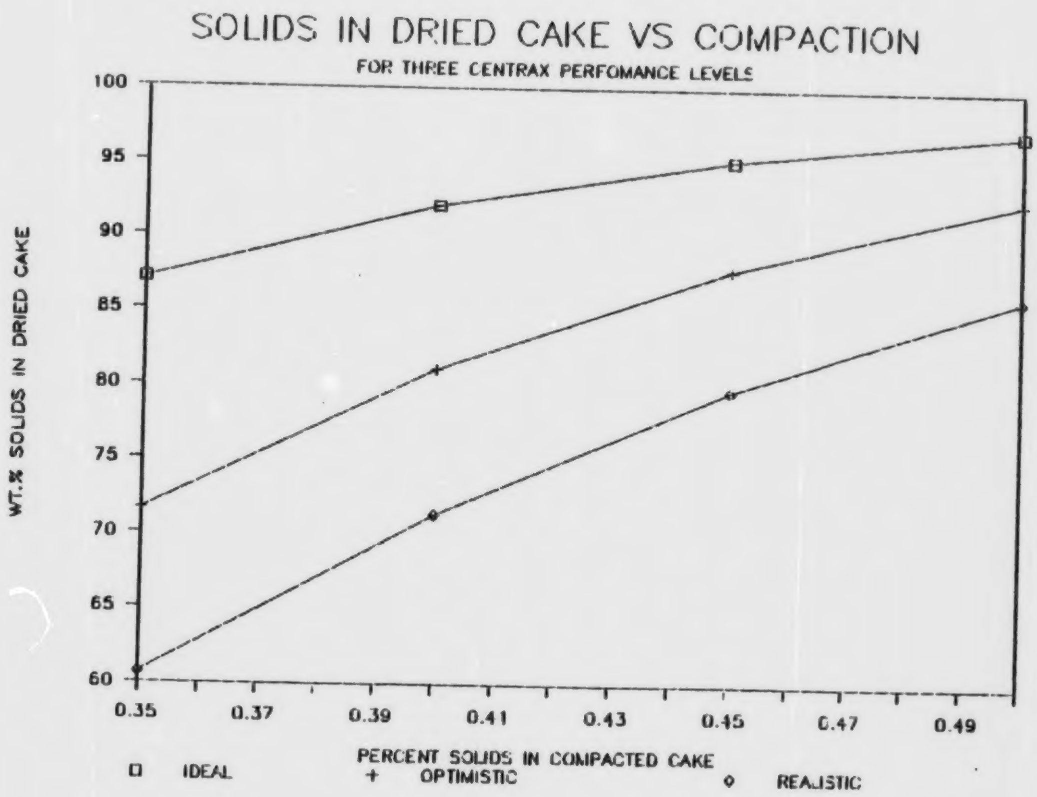


FIGURE 5-8
SOLIDS IN DRIED CAKE VS COMPACTION

PROPRIETARY INFORMATION

Results are shown for the ideal, optimistic and realistic cases discussed previously.

The degree of compaction also influences the number of CENTRAX units required to process a given quantity of deasher flow. This effect is shown in Figure 5-9 . More units are required at lower levels of cake compaction because the cake thickness permitted in the CENTRAX bowl is only 5 mm. At low compactions, more frequent feed flow interuptions are necessary for the scraper cycle.

The assumed disposition of UC in the overflow also influences the number of CENTRAX units. Figure 5-9 shows results of the analysis for the case where UC in the overflow converts and for the case where UC accumulates in the overflow without converting until a 60 percent reject is sufficient to remove the steady state accumulation. In the example shown, the number of units required is reduced by about 10 percent when UC is assumed to convert.

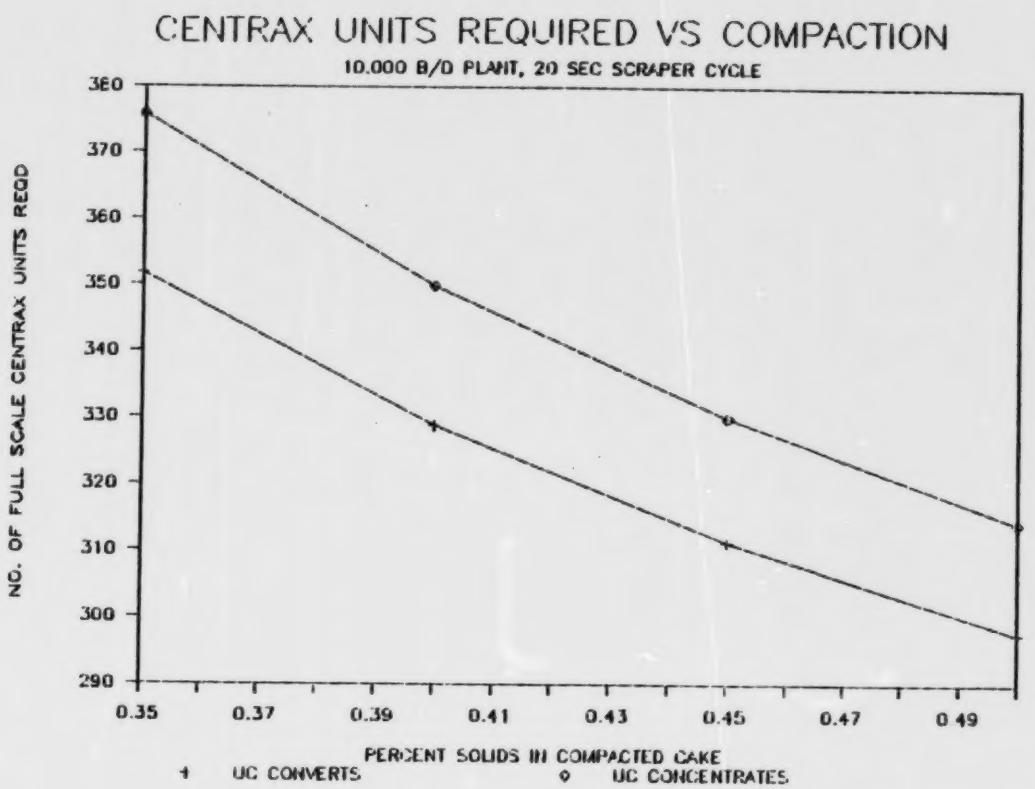


FIGURE 5-9
CENTRAX UNITS REQUIRED VS COMPACTION

PROPRIETARY INFORMATION

6.0 IMPACT OF CENTRAX DEASHING ON TWO-STAGE LIQUEFACTION ECONOMICS

6.1 Methodology

The potential economic impact of CENTRAX deashing in the context of Two-Stage direct coal liquefaction has been investigated by comparing the costs of liquid products from conceptual commercial scale plants using CENTRAX deashing to those using Kerr-McGee CSD deashing. The methodology of cost estimation used in this study is similar to that documented in an earlier MITRE study on liquefaction economics (MITRE, El Sawy et al 1986).

Briefly, a conceptual commercial design of a two-stage liquefaction plant developed for the U.S. Department of Energy by UOP/SDC is used as the baseline in the present study (UOP/SDC, 1981). When other plant configurations are evaluated, the required size of various components are compared to the baseline. Plant construction costs are determined from the baseline by using 0.7 exponent scaling based on capacity. The material flow rates through each unit operation of the conceptual commercial plant are estimated by scaling up pilot plant data to a plant size processing 30,000 tons per day of moisture-free coal to liquefaction. In this study, pilot plant operating data from the Wilsonville coal liquefaction facility were used.

The technology of two-stage liquefaction has made significant progress since the CENTRAX process was initially conceived. The feedstocks used in the CENTRAX development effort were produced using the ITSL mode where the thermal stage output was deashed before hydrotreatment in stage two. Since then it has been demonstrated that better overall liquefaction performance and product quality can be achieved if the thermally produced material is hydrotreated before deashing. In this Reconfigured Integrated Two-Stage Liquefaction configuration (RITSL) deashing occurs after hydrotreatment. Deashing is still important, but is easier because the soluble material contains fewer asphaltenes and preasphaltenes.

In recent tests, Wilsonville has used a process configuration in which the majority of the high pressure separator bottoms from the hydrotreater are recycled directly to the thermal stage without deashing. This is the ash-recycle mode of operation. In this mode, a small portion of the hydrotreater effluent is vacuum distilled and these bottoms are deashed to remove the steady-state accumulation of ash and unconverted coal.

In this economic evaluation, MITRE has compared CENTRAX and Kerr-McGee CSD deashing systems for all three configurations described above. The feedstock used in all cases was Illinois No. 6 bituminous coal.

Table 6-1 shows the Wilsonville pilot plant data used in these analyses. Note that some of the runs include a net production of residual material while others show over conversion resulting in a negative residual make. In the MITRE analysis, the space velocity of the catalytic hydrotreater is adjusted to increase or reduce resid conversion until net resid production is zero. This feature of the analysis is also used when changes in deashing performance are considered. If the deashing system results in a decrease in the residual material rejected with the solids,

TABLE 6-1
WILSONVILLE PILOT PLANT DATA USED IN ANALYSIS

ITSL

RUN 244B	TEST DATA #/# MAF
H2 STAGE 1	0.0122
H2 STAGE 2	0.0352
HETERO	0.1390
C1-C3	0.0500
NAPHTHA (350F-)	0.0965
MD (350-450F)	0.0730
MHD (450-650F)	0.2103
HD (650-850F)	0.1756
RESID (850+)	0.1066
SOL. REJECT	0.0922
UNCONVERTED COAL	0.1042
ASH	0.1172
	1.0000

RITSL

RUN 250D	TEST DATA #/# MAF
H2 STAGE 1	0.0166
H2 STAGE 2	0.0444
HETERO	0.1476
C1-C3	0.0698
NAPHTHA (350F-)	0.1330
MD (350-450F)	0.0708
MHD (450-650F)	0.2130
HD (650-850F)	0.2216
RESID (850+)	-0.0128
SOL. REJECT	0.1294
UNCONVERTED COAL	0.0886
ASH	0.1300
	1.0000

RITSL (Ash Recycle)

RUN 250G	TEST DATA #/# MAF
H2 STAGE 1	0.0185
H2 STAGE 2	0.0457
HETERO	0.1569
C1-C3	0.0767
NAPHTHA (350F-)	0.1534
MD (350-450F)	0.0775
MHD (450-650F)	0.2359
HD (650-850F)	0.1621
RESID (850+)	0.0543
SOL. REJECT	0.0818
UNCONVERTED COAL	0.0655
ASH	0.1158

then the hydrotreater is scaled accordingly to convert the increased residual to distillate.

6.2 Deasher Cost Assumptions

The installed cost of CSD deashing for a 30,000 ton per day commercial liquefaction plant using on ITSL configuration was estimated by MITRE to be 255 million dollars (1981), (MITRE El Sawy et al, 1986). The deasher flowrate or capacity, net of deashing solvent, was 44,580 tons per day or 3,715,000 lbs per hour. Deasher costs for other CSD deasher flow rates have been estimated by assuming 0.7 exponent scaling based on capacity.

In the 10,000 barrel per day conceptual commercial plant designed by Sandwell in their Phase III report, deasher costs were estimated at $3700 per ton per day of deasher feed for CSD deashing and $9200 per ton per day for CENTRAX deashing. These capacities are believed to include solvent for CSD and extractant for CENTRAX.

The total installed cost of CENTRAX deashing was estimated by Sandwell to be $46 million for a deasher flow of 5064 tons per day. The cost of the commercially scaled CENTRAX units was estimated to be $100,000 (US) per unit by Sandwell and DSMA. For the 350 units estimated by MITRE to be required for the Sandwell 10,000 barrel per day plant (See Section 5) this would be equivalent to $35 million. The remaining $11 million is assumed to be required for extractant recovery equipment.

MITRE has used the following expression to estimate the total cost of CENTRAX deashing:

$$\left[\frac{\text{Number of units required}}{(350)}\right]^{0.9} \times \$35 \text{ million}$$

$$+$$

$$\left[\frac{\text{Deasher Flow (lb/hr)}}{(3,715,000)}\right]^{0.7} \times \$11 \text{ million}$$

The 0.9 exponent factor used to scale the actual CENTRAX centrifuge units is based on an expected cost reduction for purchase of a large number of units. The 0.7 exponent factor for the extractant recovery is used since this equipment is based on total unit capacity.

Personal communications with Sharples Inc., the U. S. centrifuge Manufacturing company, have been made to obtain costs of comparable capacity vertical scroll centrifuge units. Costs quoted by them indicate that the Sandwell estimate of $100,000 per unit may be too low by a factor of 1.5 (R. Moll, Sharples, Inc., Pennsylvania, 1986). To make on allowance for this, MITRE has investigated the sensitivity of CENTRAX cost to the cost of liquid products by using the 1.5 factor, suggested by Sharples, in addition to the base cost suggested by Sandwell. All MITRE economic analyses use both cost factors for CENTRAX deashing.

6.3 Economic Comparisons of CENTRAX and Kerr-McGee CSD in the ITSL Mode:

Wilsonville run 244 B is an example of operations performed in the Integrated Two-Stage Liquefaction (ITSL) mode using Illinois #6 coal.

Figure 6-1 illustrates the Wilsonville test data for this run based on 100 lbs of moisture-free coal as input. The CSD deasher produces an ash concentrate containing 70.6 weight percent solids.

Table 6-2 is a summary of the economic performance of Wilsonville 244B using CSD deashing. This table is the output of the MITRE coal liquefaction model that details the results of scaling the test data to a conceptual commercial plant processing 30,000 tons per day of moisture-free coal to the liquefaction reactors. The yield structure, after modification to allow for net zero resid make, is shown on the left of Table 6-2 based on wt-percent MAF coal. The scaled output from this plant is shown as barrels per day of raw and hydrotreated products in the center of the table. The average quality of the products is shown based on elemental composition and API (American Petroleum Institute) gravity for both raw and hydrotreated liquids.

The column on the right of Table 6-2 shows the coal inputs to this conceptual plant. In addition to liquefaction, coal is also required for steam generation and for hydrogen production. Plant construction costs disagregated by plant area is also shown. The total construction cost is then converted into total capital required by the procedure outlined in an earlier MITRE report. (MITRE, El Sawy et al, 1986). This entails additions of royalties, allowance for funds used during construction, start-up costs, working capital and initial charge of catalyst and chemicals.

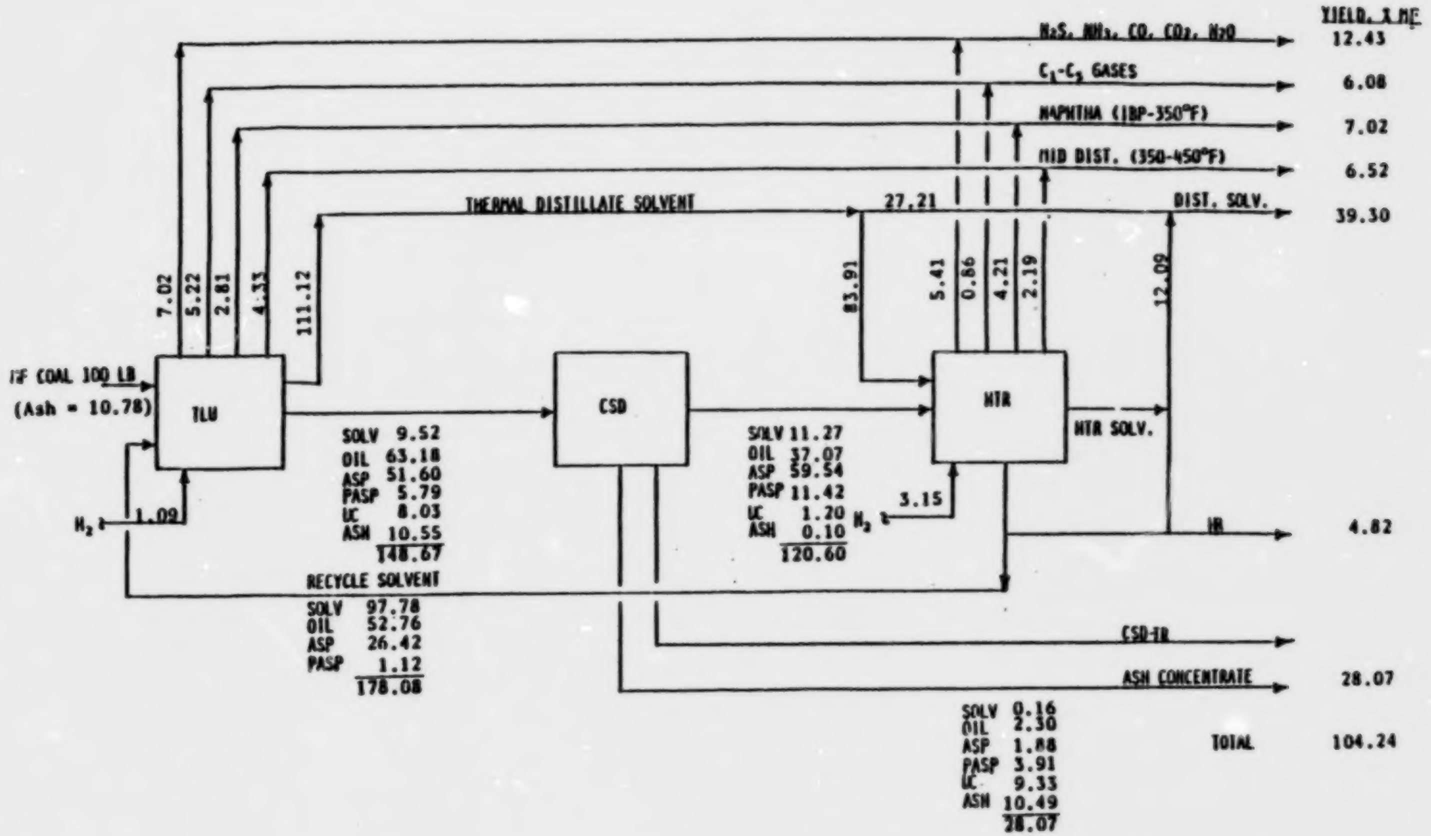

YIELD, % MF
H₂S, NH₃, CO, CO₂, H₂O
12.43
C₁-C₃ GASES
6.08
NAPHTHA (IBP-350°F)
7.02
MID DIST. (350-450°F)
6.52
THERMAL DISTILLATE SOLVENT
27.21
DIST. SOLV.
39.30
7.02
5.22
2.81
4.33
111.12
83.91
5.41
0.86
4.21
2.19
12.09
IF COAL 100 LB
(Ash = 10.78)
TLU
CSD
HTR
HTR SOLV.
H₂
1.09
H₂
3.15
SOLV 9.52
OIL 63.18
ASP 51.60
PASP 5.79
UC 8.03
ASH 10.55
148.67
SOLV 11.27
OIL 37.07
ASP 59.54
PASP 11.42
UC 1.20
ASH 0.10
120.60
HR
4.82
RECYCLE SOLVENT
SOLV 97.78
OIL 52.76
ASP 26.42
PASP 1.12
178.08
CSD-IR
ASH CONCENTRATE
28.07
SOLV 0.16
OIL 2.30
ASP 1.88
PASP 3.91
UC 9.33
ASH 10.49
28.07
TOTAL
104.24

FIGURE 6-1
MATERIAL BALANCE FLOW DIAGRAM FOR RUN 244B

PROPRIETARY INFORMATION

TABLE 6-2
TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-244B
ILLINOIS #6 RUN TS244B, K-M CRITICAL SOLVENT DEASHING

```
                        TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    244B             70.6000 SOLID REJECT
                                        ILLINOIS # 6 RUN TS244B, K-M CRITICAL SOLVENT DEASHING

                                                    COMMERCIAL PLANT OUTPUT             COAL TO LIQUEFACTION, T/D AR      33515
 YIELD STRUCTURE,             WT% MAF COAL          LBS/HR      BBLS/DAY                COAL TO STEAM,        T/D AR       1996
                                                                                        COAL TO HYDROGEN,     T/D AR       3522
 H2 Consumed, Stage 1          0.0122                                                       TOTAL COAL        T/D AR      39033
 H2 Consumed, Stage 2          0.0389
 Heteroatoms (inc H2O)         0.1397                                                   PLANT CONSTRUCTION COST ($1000)
 C1C3 Gases                    0.0526                                                       LIQUEFACTION                1476595
 C4 - 350F                     0.0965              215942       18253                       SOLIDS REMOVAL               278959
 350F - 450F Product           0.0730              163355       11870                       HYDROGEN PRODUCTION          578755
 450F - 650F Product           0.2103              470596       34195                       BALANCE OF PLANT             443473
 650 - 850F Product            0.2826              632449       42494                         TOTAL CONS COST           2777782
    TOTAL C4-850F PRODUCT      0.6624
 850F+ Product                 0.0000                                                   TOTAL CAPITAL REQUIRED          4457159
 Soluble Reject                0.0922
 Unconverted Coal              0.1042                                                   OPERATING COSTS,$1000/YR
 Ash                           0.1172                                                       COAL COST                    386424
     TOTAL,RAW PRODUCT                            1482341      106812                       OTHER OPERATING COSTS        385811
     TOTAL, HYDROTREATED PRODUCT                  1476232      122675                       BY-PRODUCT CREDITS            39592
                                                                                              TOTAL,$1000/YR             732642
 AVERAGE PRODUCT QUALITY
                      CARBON   HYDROGEN     NITROGEN     SULFUR       OXYGEN      API deg       HYDROTREATING COSTS          329748
RAW                    87.34      10.60         0.24       0.21         1.60        17.35         TOTAL INC.HYDROTREAT      1062391
HYDR                   87.00      13.00                                             40.00
                                                                                        REQUIRED SELLING PRICE            $/BBL
 DEASHER COST FACTOR            1.00
 COAL CLEANING, %Btu Recovered  100.00 %         0 Flag,1 or 0                             RAW PRODUCT                    38.11
 COAL DRYING FLAG (1 or 0)         0                                                        HYDROTREATED PRODUCT           41.33


                                            SG          C            H          N          S          O         API

                             NAP          0.8104     85.2100      12.8600     0.0845     0.3600     1.5000     43.11
                             MD           0.9427     86.3400      10.7300     0.2300     0.2200     2.4800     18.60
                             HMD          0.9427     86.3400      10.7300     0.2300     0.2200     2.4800     18.60
                             PHD          1.0195     89.0700       9.6900     0.3100     0.1600     0.7600      7.29
```

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Operating costs are shown for coal (charged at $30 per ton on a dry basis) and other operating costs. These include electric power, catalysts and chemicals, water, operating labor, administration, maintenance, local taxes and insurance and solids disposal costs. By-product credits are given for sulfur ($50 per ton) and ammonia ($100 per ton). Costs of hydrotreatment for the raw products are included when the costs of hydrotreated products are computed. The cost is assumed to be proportional to the quantity of hydrogen required to convert the raw products to a 13 weight percent hydrogen, 40 API product containing no heteroatoms. The hydrogen required is computed from mass balance under the assumption that all oxygen, sulphur and nitrogen, and 1 percent of carbon react with hydrogen and are evolved as gases. In these runs, the cost of hydrotreatment is assumed to be $1/lb. of hydrogen required. This estimate is representative of operating costs for hydrotreatment in existing refineries.

The required selling price in dollars per barrel is then calculated from the total capital and operating costs assuming a standard set of economic assumptions. These are shown in Table 6-3 . This procedure of using a constant methodology and standard economic assumptions for the analysis of all the cases assures that the difference between costs is meaningful.

The economics of the ITSL plant using the CSD deashing process provides the baseline performance and economics against which the CENTRAX cases are compared. MITRE has developed conceptual commercial plants

TABLE 6-3
ECONOMIC ANALYSIS ASSUMPTIONS

All analysis assumptions are May 1981 dollars.

A five-year construction period is assumed.

20-year plant life

16-year tax life (DDB)

75/25 debt equity

17-percent interest

10-percent inflation rate

DCF = 20 percent (20-percent ROI)

20-percent investment tax credit*

46-percent income tax rate

---

*Assumed to be taken in year 5 to approximate UOP/SDC assumptions.

based on the projected use of CENTRAX deashing in the ITSL configuration using run 244B as the basis of plant performance. Three assumptions of CENTRAX performance level are analyzed: ideal, optimistic and realistic.

The IDEAL performance assumes that the unconverted coal in the overflow converts on recycle and that all extractant and non-distillate oils (maltenes) can be recovered from the compacted cake during the drying process. This case further assumes that there is no preferential precipitation of A and PA during the centrifugation process. Thus the compacted cake contains the theoretical concentration of these materials under perfect mixing assumptions as predicted in the MITRE CENTRAX Simulation model.

The OPTIMISTIC performance level assumes that the UC in the overflow does not convert but builds up until steady-state accumulation is removed. It further assumes that the maltenes cannot be recovered during drying. In the REALISTIC case. it is assumed that the A and PA concentrate to twice the level of that predicted by perfect mixing and that in addition to maltenes, 5 percent of distillable oil remains in the underflow after drying.

Figure 6-2 illustrates the composition of the dried compacted cake for these three sets of assumptions. Solids content drops from 96.4 weight percent for the ideal case to 81.8 percent for the realistic.

Tables 6-4, 6-5, and 6-6 show the results of the MITRE economic analysis for the three sets of assumptions described above using CENTRAX

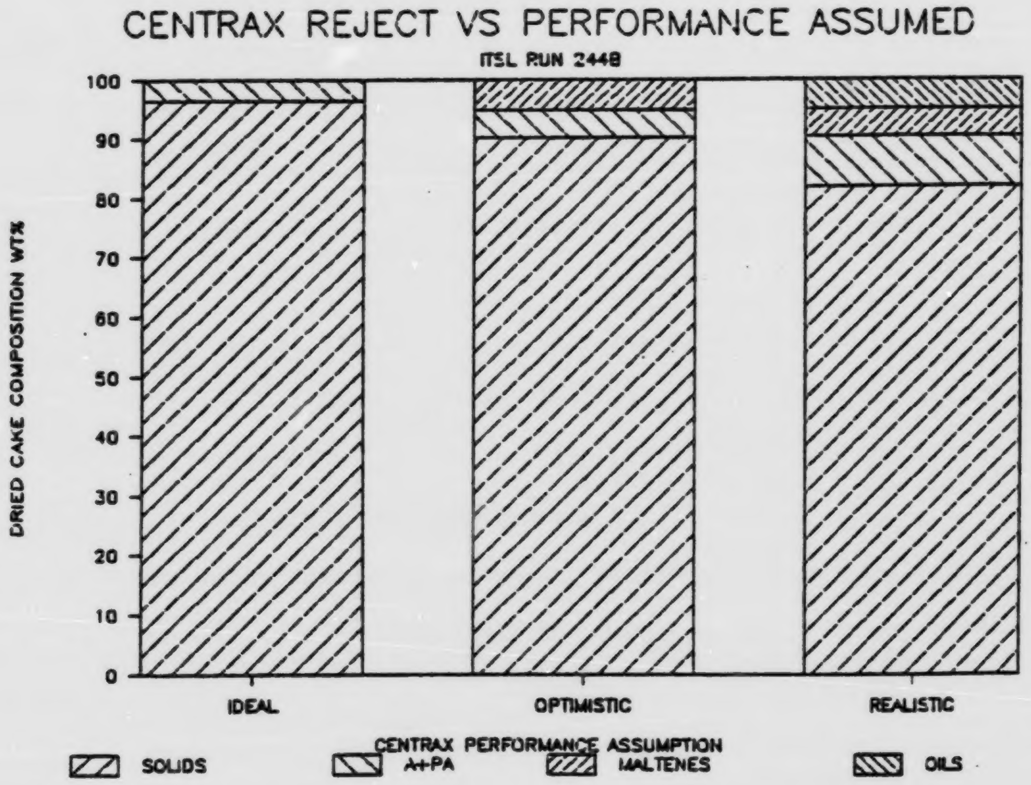

FIGURE 6-2
CENTRAX REJECT VS PERFORMANCE ASSUMED

TABLE 6-4
TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-244B
ILLINOIS #6 RUN TS244B CENTRAX DEASHING

```
                  TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    244B          96.4100 SOLID REJECT
                               ILLINOIS # 6 RUN TS244B CENTRAX DEASHING

                                                     COMMERCIAL PLANT OUTPUT                   COAL TO LIQUEFACTION, T/D AR     33515
  YIELD STRUCTURE,              #/# MAF COAL           LBS/HR       BBLS/DAY                   COAL TO STEAM,        T/D AR      2172
                                                                                               COAL TO HYDROGEN,     T/D AR      9149
  H2 Consumed, Stage 1            0.0129                                                           TOTAL COAL        T/D AR     44836
  H2 Consumed, Stage 2            0.0443
  Heteroatoms (inc H2O)           0.1482                                                       PLANT CONSTRUCTION COST ($1000)
  C1C3 Gases                      0.0577                                                           LIQUEFACTION                1553045
  C4 - 350F                       0.1019              228067        19278                          SOLIDS REMOVAL               480496
  350F - 450F Product             0.0771              172527        12536                          HYDROGEN PRODUCTION          612855
  450F - 650F Product             0.2221              497021        36115                          BALANCE OF PLANT             469147
  650 - 850F Product              0.3895              871614        58564                            TOTAL CONS COST           3115543
     TOTAL C4-850F PRODUCT        0.7906
  850F+ Product                   0.0000                                                       TOTAL CAPITAL REQUIRED          4982841
  Soluble Reject                  0.0064
  Unconverted Coal                0.0539                                                       OPERATING COSTS,$1000/YR
  Ash                             0.1172                                                           COAL COST                    443872
     TOTAL,RAW PRODUCT                               1769229       126493                          OTHER OPERATING COSTS        399759
     TOTAL, HYDROTREATED PRODUCT                     1765954       146751                          BY-PRODUCT CREDITS            63566
                                                                                                     TOTAL,$1000/YR             780066
 AVERAGE PRODUCT QUALITY
                      CARBON   HYDROGEN      NITROGEN     SULFUR       OXYGEN      API deg         HYDROTREATING COSTS          411063
RAW                    87.54      10.49          0.25       0.21         1.51        16.19           TOTAL INC.HYDROTREAT      1191129
HYDR                   87.00      13.00                                              40.00
                                                                                               REQUIRED SELLING PRICE            $/BBL
  DEASHER COST FACTOR             1.50
  COAL CLEANING, %Btu Recovered  100.00 %          0 Flag,1 or 0                                   RAW PRODUCT                   35.04
  COAL DRYING FLAG (1 or 0)          0                                                             HYDROTREATED PRODUCT          38.69


                                          SG            C             H          N          S          O          API

                            MAF         0.8104       85.2100       12.8600    0.0845     0.3600     1.5000       43.11
                            HD          0.9427       86.3400       10.7300    0.2300     0.2200     2.4800       18.60
                            HHD         0.9427       86.3400       10.7300    0.2300     0.2200     2.4800       18.60
                            PHD         1.0195       89.0700        9.6900    0.3100     0.1600     0.7600        7.29
```

PROPRIETARY INFORMATION

TABLE 6-5
TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-244B
ILLINOIS #6 RUN TS244B CENTRAX DEASHING

```
                    TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    244B            90.1700 SOLID REJECT
                                  ILLINOIS # 6 RUN TS244B CENTRAX DEASHING

                                                     COMMERCIAL PLANT OUTPUT                 COAL TO LIQUEFACTION, T/D AR      33515
 YIELD STRUCTURE,              WT% MAF COAL           LBS/HR      BBLS/DAY                 . COAL TO STEAM,         T/D AR       2172
                                                                                             COAL TO HYDROGEN,      T/D AR       6914
 H2 Consumed, Stage 1              0.0124                                                        TOTAL COAL         T/D AR      42601
 H2 Consumed, Stage 2              0.0420
 Heteroatoms (inc H2O)             0.1425                                                    PLANT CONSTRUCTION COST ($1000)
 C1C3 Gases                        0.0551                                                        LIQUEFACTION                 1533428
 C4 - 350F Product                 0.0981             219413        18546                        SOLIDS REMOVAL                509528
 350F - 450F Product               0.0742             165981        12061                        HYDROGEN PRODUCTION           595296
 450F - 650F Product               0.2137             478161        34745                        BALANCE OF PLANT              463595
 650 - 850F Product                0.3586             802395        53913                          TOTAL CONS COST            3101848
    TOTAL C4-850F PRODUCT          0.7445
 850F+ Product                     0.0000                                                    TOTAL CAPITAL REQUIRED           4952528


 Ash                               0.1172                                                        COAL COST                     421747
    TOTAL,RAW PRODUCT                                1665949       119265                        OTHER OPERATING COSTS         398993
    TOTAL, HYDROTREATED PRODUCT                      1662235       130132                        BY-PRODUCT CREDITS             52569
                                                                                                   TOTAL,$1000/YR              768171
 AVERAGE PRODUCT QUALITY
                      CARBON    HYDROGEN      NITROGEN    SULFUR      OXYGEN      API deg        HYDROTREATING COSTS           384319
RAW                    87.51      10.51          0.25      0.21        1.52        16.38           TOTAL INC.HYDROTREAT       1152490
HYDR                   87.00      13.00                                            40.00
                                                                                             REQUIRED SELLING PRICE            $/BBL
 DEASHER COST FACTOR               1.50
 COAL CLEANING, %Btu Recovered   100.00 %         0 Flag,1 or 0                                  RAW PRODUCT                    36.76
 COAL DRYING FLAG (1 or 0)            0                                                          HYDROTREATED PRODUCT           40.17



                                     SG         C           H          N        S          O          API

                         NAP       0.8104     85.2100     12.8600    0.0845   0.3600     1.5000      43.11
                         RD        0.9427     86.3400     10.7300    0.2300   0.2200     2.4000      18.60
                         HHD       0.9427     86.3400     10.7300    0.2300   0.2200     2.4000      18.60
                         PHD       1.0195     89.0700      9.6900    0.3100   0.1600     0.7600       7.29
```

PROPRIETARY INFORMATION

TABLE 6-6
TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-244B
ILLINOIS #6 RUN TS244B CENTRAX DEASHING

```
                      TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    244B            81.8500 SOLID REJECT
                                   ILLINOIS # 6 RUN TS244B CENTRAX DEASHING

                                                      COMMERCIAL PLANT OUTPUT               COAL TO LIQUEFACTION, T/D AR     33515
  YIELD STRUCTURE,              WT% MAF COAL            LBS/HR     BBLS/DAY                 COAL TO STEAM,        T/D AR      2172
                                                                                            COAL TO HYDROGEN,     T/D AR      5992
  H2 Consumed, Stage 1            0.0124                                                       TOTAL COAL         T/D AR     41679
  H2 Consumed, Stage 2            0.0412
  Heteroatoms (inc H2O)           0.1423                                                    PLANT CONSTRUCTION COST ($1000)
  C1C3 Gases                      0.0546                                                       LIQUEFACTION                1515491
  C4 - 350F Product               0.0981                219413      18546                      SOLIDS REMOVAL               509528
  350F - 450F Product             0.0742                165981      12061                      HYDROGEN PRODUCTION          590673
  450F - 650F Product             0.2137                478161      34745                      BALANCE OF PLANT             460460
  650 - 850F Product              0.3351                749968      50390                        TOTAL CONS COST           3076152
     TOTAL C4-850F PRODUCT        0.7211
  850F+ Product                   0.0000                                                    TOTAL CAPITAL REQUIRED         4909994
  Soluble Reject                  0.0459
  Unconverted Coal                0.0898                                                    OPERATING COSTS,$1000/YR
  Ash                             0.1172                                                       COAL COST                    412620
      TOTAL,RAW PRODUCT                                1613522     115742                      OTHER OPERATING COSTS        397039
      TOTAL, HYDROTREATED PRODUCT                      1608990     133707                      BY-PRODUCT CREDITS            49469
                                                                                                  TOTAL,$1000/YR            760190
 AVERAGE PRODUCT QUALITY
                        CARBON   HYDROGEN       NITROGEN      SULFUR      OXYGEN      API deg      HYDROTREATING COSTS          368153
RAW                      87.46      10.54           0.25        0.21        1.55        16.68         TOTAL INC.HYDROTREAT     1128343
HYDR                     87.00      13.00                                               40.00
                                                                                             REQUIRED SELLING PRICE          $/BBL
  DEASHER COST FACTOR               1.50
  COAL CLEANING, %Btu Recovered   100.00 %        0 Flag,1 or 0                                RAW PRODUCT                   37.51
  COAL DRYING FLAG (1 or 0)            0                                                       HYDROTREATED PRODUCT          40.82




                                                    SG          C           H          N          S          O        API

                                       NAP      0.8104     85.2100     12.8600     0.0845     0.3600     1.5000     43.11
                                       HD       0.9427     86.3400     10.7300     0.2300     0.2200     2.4800     18.60
                                       NHD      0.9427     86.3400     10.7300     0.2300     0.2200     2.4800     18.60
                                       PHD      1.0195     89.0700      9.6900     0.3100     0.1600     0.7600      7.29
```

PROPRIETARY INFORMATION

deashing. Results are shown here for the deasher cost factor of 1.5. Tables for the DAF of 1.0 are shown in Appendix A.

The cost of liquid products using CENTRAX deashing for the ideal, optimistic and realistic cases compared to CSD is illustrated graphically in Figure 6-3 for the ITSL mode using run 244B as the basis. The number of CENTRAX Units (actual centrifuges) necessary are 4132 for the ideal case and 4459 for both the optimistic and realistic cases. Both raw and hydrotreated product costs are shown for both CSD and CENTRAX. The two cost factors for CENTRAX deashing are also shown. For the hydrotreated products, CSD cost is $41.33 per barrel. This is compared to $38.69, $40.17 and $40.82 per barrel for CENTRAX in going from ideal to realistic. The cost differential between realistic CENTRAX and CSD is thus only about 1 percent; which is essentially an economic stand off.

6.4 Economic Comparisons of CENTRAX and CSD in the RITSL Mode:

Figure 6-4 shows the Wilsonville test data for run 250D performed in the RITSL configuration. CSD operation produces an ash concentrate containing 62.8 weight percent solids. The results of the MITRE economic model applied to this set of test data are shown in Table 6-7.

The identical analytical procedures applied to the ITSL mode in Section 6.3 are applied here. CSD deashing is replaced by CENTRAX for each of the three assumptions described above. Figure 6-5 shows the dried cake composition for each of these cases.

Application of the MITRE economic model for coal liquefaction for each of these CENTRAX assumptions results in liquid product costs for the

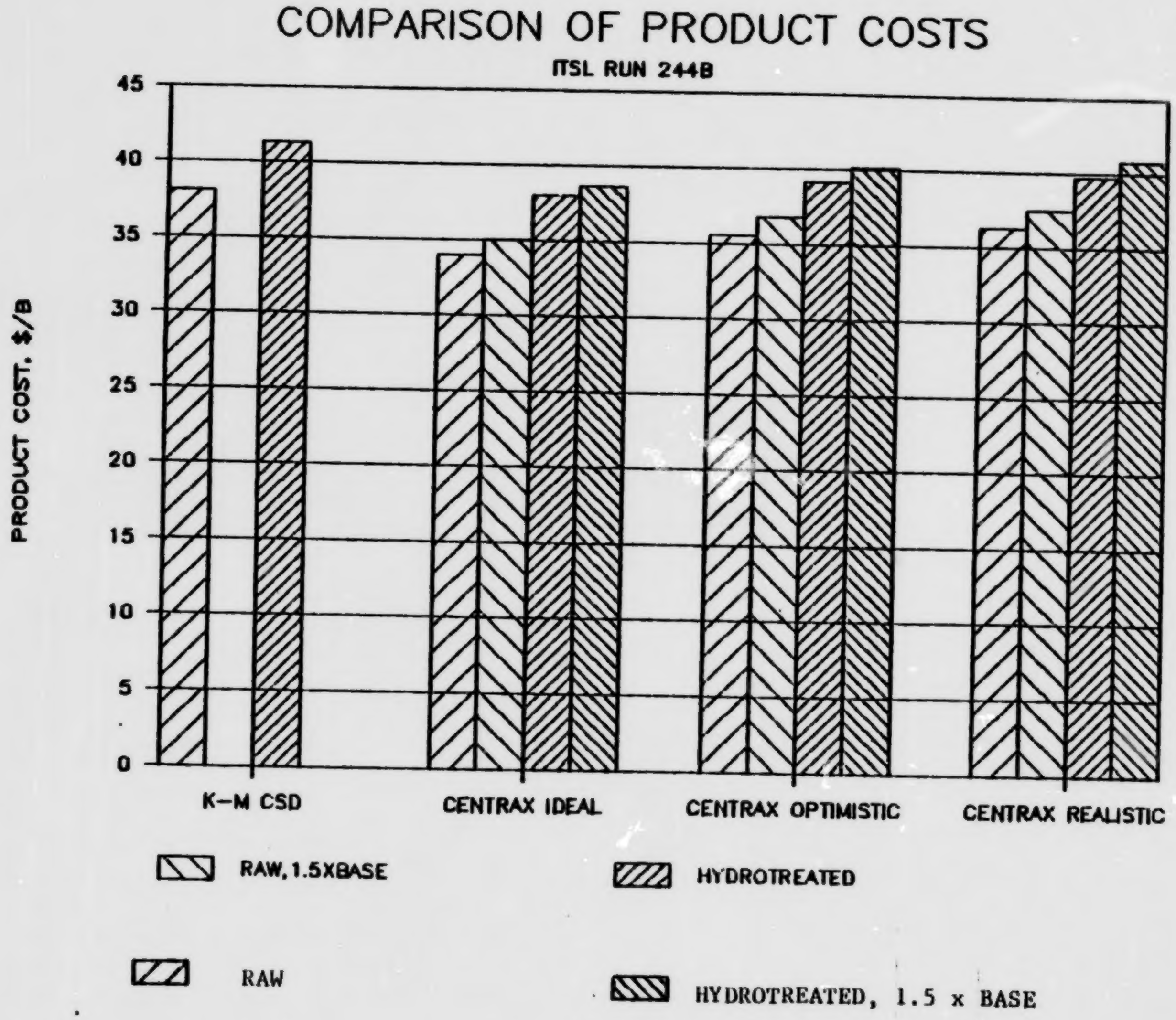

FIGURE 6-3
COMPARISON OF PRODUCT COSTS

PROPRIETARY INFORMATION

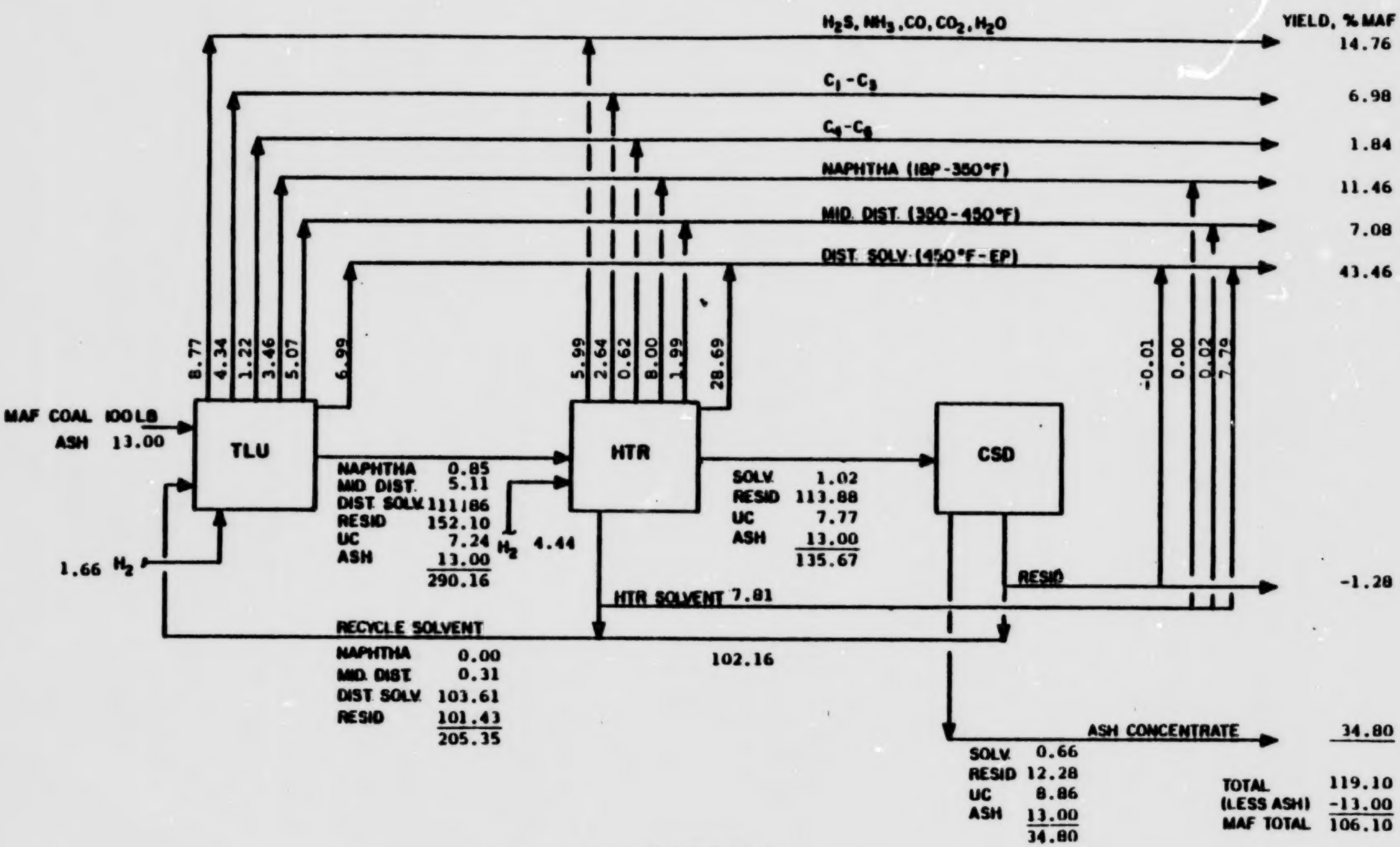
H$_2$S, NH$_3$, CO, CO$_2$, H$_2$O
YIELD, % MAF
14.76
C$_1$-C$_3$
6.98
C$_4$-C$_5$
1.84
NAPHTHA (IBP-350°F)
11.46
MID. DIST. (350-450°F)
7.08
DIST. SOLV. (450°F-EP)
43.46
8.77
4.34
1.22
3.46
5.07
6.99
5.99
2.64
0.62
8.00
1.99
28.69
-0.01
0.00
0.02
7.79
MAF COAL 100 LB
ASH 13.00
TLU
HTR
CSD
1.66 H$_2$
H$_2$ 4.44
NAPHTHA 0.85
MID. DIST. 5.11
DIST. SOLV. 111.86
RESID 152.10
UC 7.24
ASH 13.00
290.16
SOLV. 1.02
RESID 113.88
UC 7.77
ASH 13.00
135.67
RESID
-1.28
HTR SOLVENT 7.81
RECYCLE SOLVENT
NAPHTHA 0.00
MID. DIST. 0.31
DIST. SOLV. 103.61
RESID 101.43
205.35
102.16
ASH CONCENTRATE
34.80
SOLV. 0.66
RESID 12.28
UC 8.86
ASH 13.00
34.80
TOTAL 119.10
(LESS ASH) -13.00
MAF TOTAL 106.10

FIGURE 6-4
MATERIAL BALANCE FLOW DIAGRAM FOR 250D

PROPRIETARY INFORMATION

TABLE 6-7
TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-250D
ILLINOIS #6 RUN TS250D, K-M CRITICAL SOLVENT DEASHING

```
              TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250D            62.0000 SOLID REJECT
                                  ILLINOIS # 6 RUN TS250D,K-M CRITICAL SOLVENT DEASHING

                                                   COMMERCIAL PLANT OUTPUT
 YIELD STRUCTURE,              WT% MAF COAL          LBS/HR      BBLS/DAY               COAL TO LIQUEFACTION, T/D AR     33135
                                                                                        COAL TO STEAM,        T/D AR      2418
 H2 Consumed, Stage 1              0.0166                                               COAL TO HYDROGEN,     T/D AR      4592
 H2 Consumed, Stage 2              0.0439                                                  TOTAL COAL         T/D AR     40145
 Heteroatoms (inc H2O)             0.1475
 C1C3 Gases                        0.0695                                               PLANT CONSTRUCTION COST ($1000)
 C4 - 350F                         0.1330             294240       25841                   LIQUEFACTION                1301028
 350F - 450F Product               0.0708             156637       12138                   SOLIDS REMOVAL               220873
 450F - 650F Product               0.2130             471239       34561                   HYDROGEN PRODUCTION          643616
 650 - 850F Product                0.2087             461636       31559                   BALANCE OF PLANT             447446
    TOTAL C4-850F PRODUCT          0.6255                                                    TOTAL CONS COST           2612964
 850F+ Product                     0.0000
 Soluable Reject                   0.1295                                               TOTAL CAPITAL REQUIRED         4207998
 Undesolved Coal                   0.0886
                                                                                        OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                                1383759      104098                   COAL COST                    397434
    TOTAL, HYDROTREATED PRODUCT                      1379577      114643                   OTHER OPERATING COSTS        356308
                                                                                           BY-PRODUCT CREDITS            85092
  AVERAGE PRODUCT QUALITY                                                                    TOTAL,$1000/YR             668650
                       CARBON   HYDROGEN     NITROGEN     SULFUR      OXYGEN      API deg
RAW                     87.44     11.79        0.23        0.10        0.44        23.90    HYDROTREATING COSTS          163036
HYDR                    87.00     13.00                                            40.00      TOTAL INC.HYDROTREAT       831686

  DEASHER COST FACTOR              1.00                                                 REQUIRED SELLING PRICE            $/BBL
  COAL CLEANING, %Btu Recovered  100.00 %          0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)          0                                                     RAW PRODUCT                   36.25
                                                                                           HYDROTREATED PRODUCT          37.22


                          SG          C           H          N          S          O         API

             NAP        0.7800      85.4100     14.2600     0.0400     0.1000     0.1100     49.91
             MD         0.8840      86.1900     12.4400     0.1300     0.1300     1.1100     28.57
             HHD        0.9340      87.8100     11.5300     0.2500     0.0800     0.3300     20.00
             PHD        1.0020      88.7700     10.2700     0.3600     0.0700     0.5300      9.72
```

PROPRIETARY INFORMATION

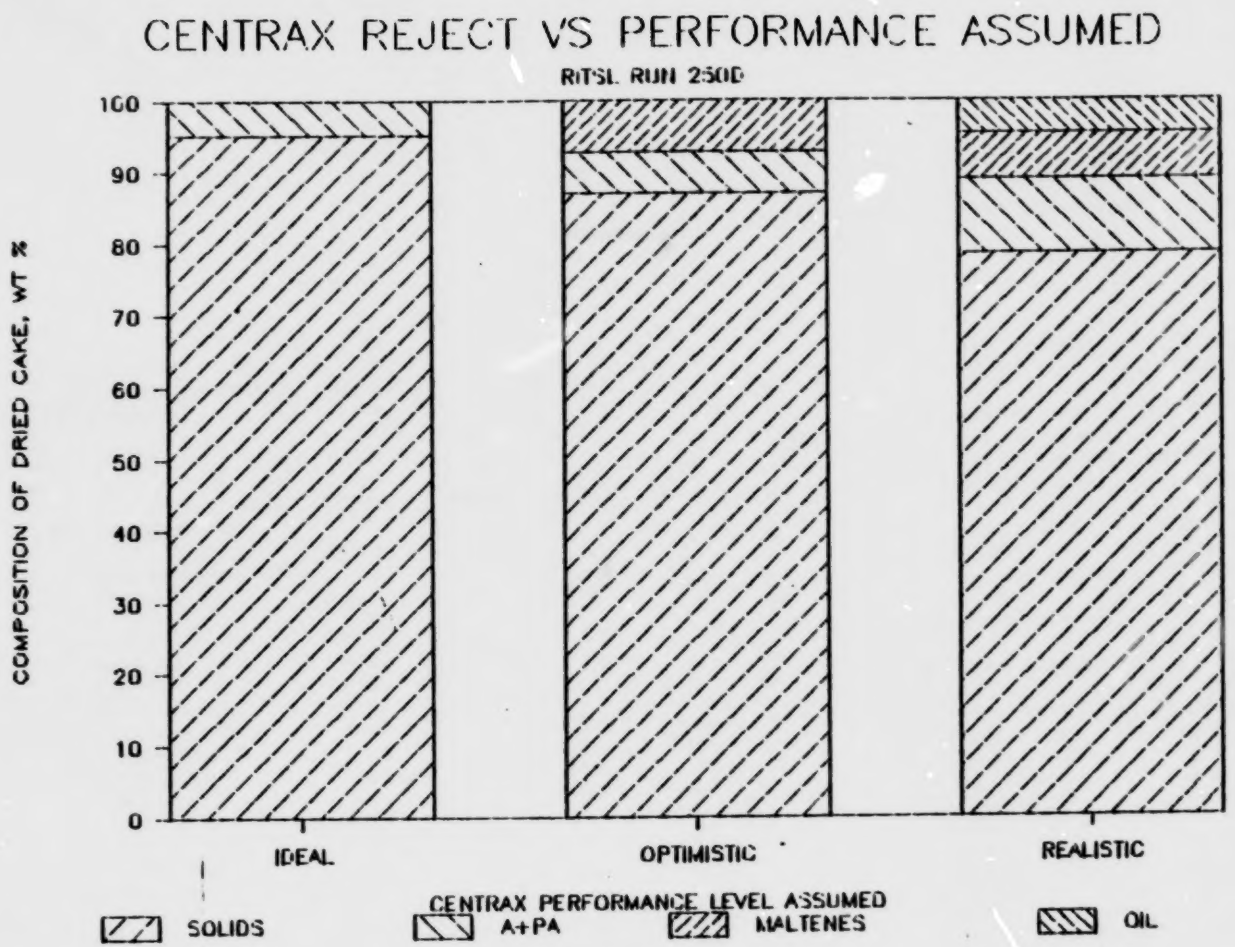

FIGURE 6-5
CENTRAX REJECT VS PERFORMANCE ASSUMED

PROPRIETARY INFORMATION

RITSL mode shown in Figure 6-6. Details of the economics for these cases are documented in Appendix A. For run 250D, CSD product costs for hydrotreated liquids are $37.22 per barrel. CENTRAX deashing costs range from $34.41 to $36.55 per barrel going from ideal to realistic assumptions. The number of CENTRAX units are 3497 for the ideal case and 3770 for the optimistic and realistic cases. Product cost savings are thus 1.8 percent using CENTRAX under realistic assumptions of compacted cake composition.

Wilsonville also operated RITSL run 250 in the ash recycle mode during period 250G. The material balance test data for this case is illustrated in Figure 6-7. Since most of the separator bottoms from the hydrotreater are recycled to the thermal unit, the material flow to the deasher is much lower in this case. This results in a lower capital cost requirement for the CSD unit. Table 6-8 shows the economic results of the ash recycle mode (run 250G) using CSD.

Figure 6-8 shows the composition of the dried cake from the CENTRAX unit when operated with deasher feed from run 250G at the 1:0.4 feed to extractant ratio recommended by Sandwell. The performance is seen to be poorer (e.g., the percent solids in the cake is lower) than in the previous cases. This is because of the reduced deasher feed and the higher solids to liquids ratio in the combined feed and extractant. The CENTRAX performance with this feed can be improved by increasing the feed to extractant ratio to 1:0.7.

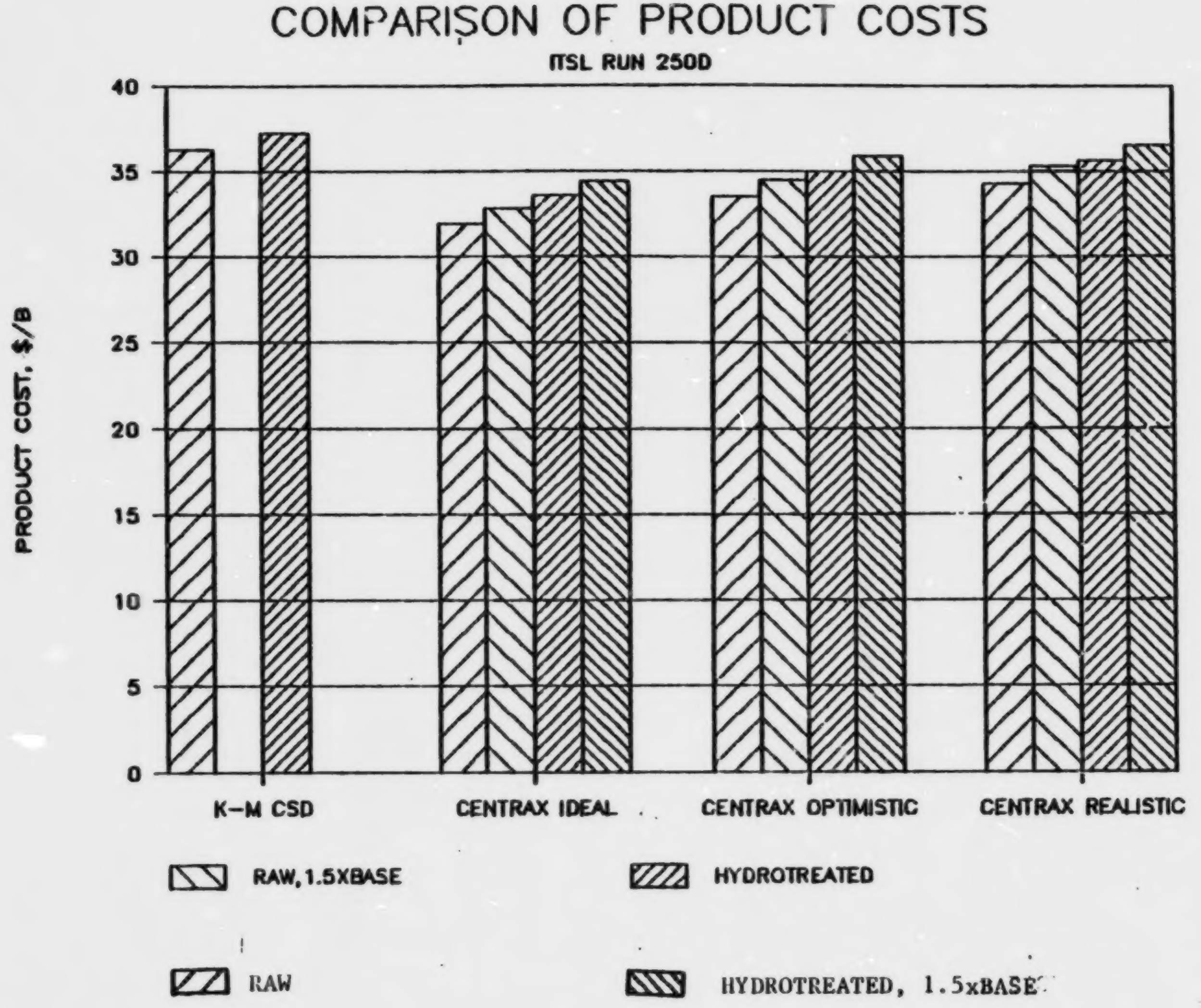

FIGURE 6-6
COMPARISON OF PRODUCT COSTS

PROPRIETARY INFORMATION

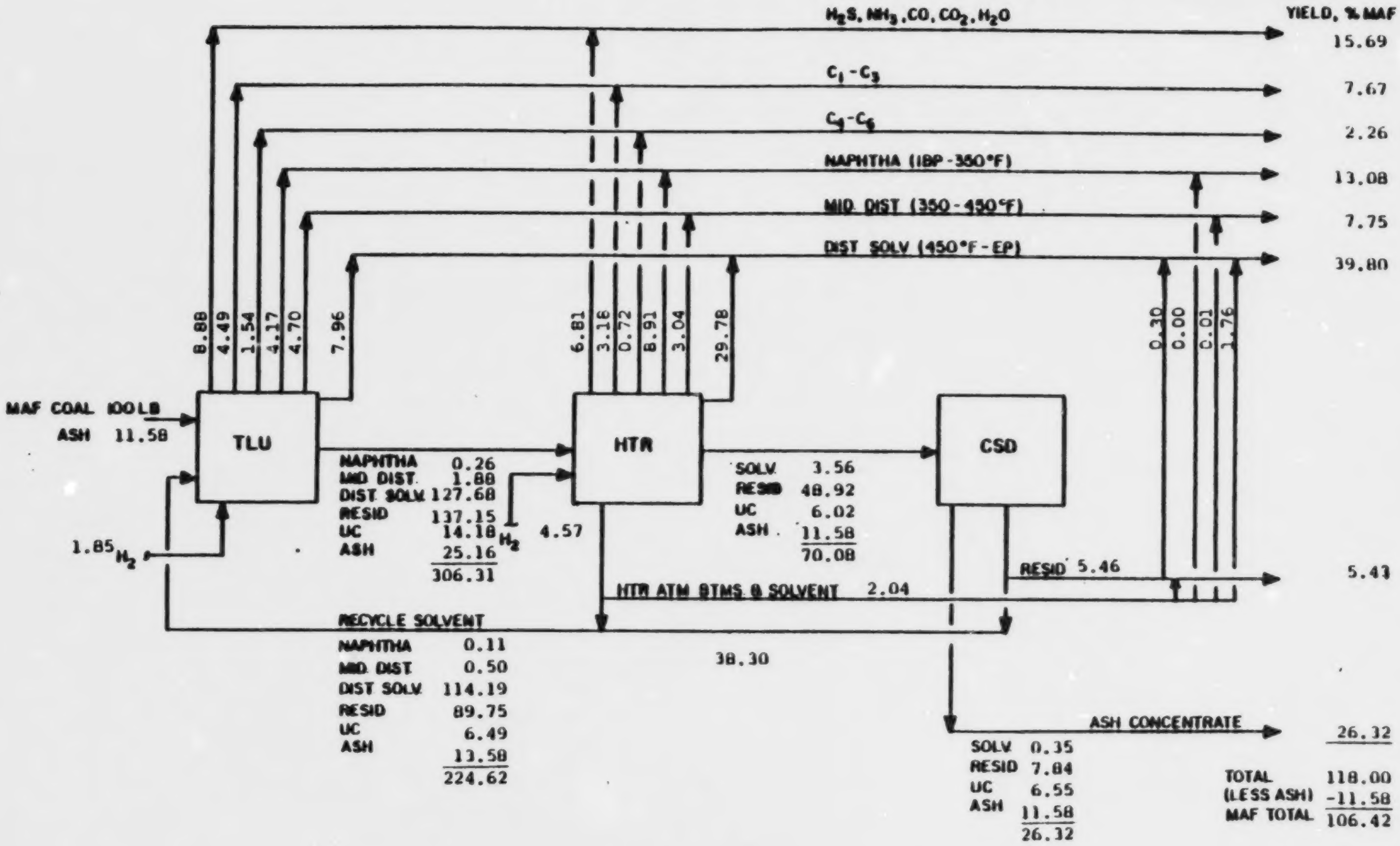
MAF COAL 100 LB
ASH 11.58
1.85 H2
TLU
HTR
CSD
8.88
4.49
1.54
4.17
4.70
7.96
6.81
3.16
0.72
8.91
3.04
29.78
0.30
0.00
0.01
1.76
H2S, NH3, CO, CO2, H2O
C1-C3
C4-C6
NAPHTHA (IBP-350°F)
MID DIST (350-450°F)
DIST SOLV (450°F-EP)
YIELD, % MAF
15.69
7.67
2.26
13.08
7.75
39.80
5.43
26.32
NAPHTHA 0.26
MID DIST. 1.88
DIST SOLV 127.68
RESID 137.15
UC 14.18
ASH 25.16
306.31
H2 4.57
SOLV 3.56
RESID 48.92
UC 6.02
ASH 11.58
70.08
RESID 5.46
HTR ATM BTMS & SOLVENT 2.04
RECYCLE SOLVENT
NAPHTHA 0.11
MID DIST 0.50
DIST SOLV 114.19
RESID 89.75
UC 6.49
ASH 13.58
224.62
38.30
ASH CONCENTRATE
SOLV 0.35
RESID 7.84
UC 6.55
ASH 11.58
26.32
TOTAL 118.00
(LESS ASH) -11.58
MAF TOTAL 106.42

FIGURE 6-7
MATERIAL BALANCE FLOW DIAGRAM FOR 250G

PROPRIETARY INFORMATION

TABLE 6-8
TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-250G
ILLINOIS #6 TS250G K-M CRITICAL SOLVENT DEASHING

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                    TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G          48.8000 SOLID REJECT
                                  ILLINOIS # 6 TS250G K-M CRITICAL SOLVENT DEASHING

                                           . COMMERCIAL PLANT OUTPUT               COAL TO LIQUEFACTION, T/D AR     33557
YIELD STRUCTURE,            WT% MAF COAL       LBS/HR      BBLS/DAY                COAL TO STEAM,        T/D AR      2651
                                                                                   COAL TO HYDROGEN,     T/D AR      8032
H2 Consumed, Stage 1           0.0185                                                 TOTAL COAL         T/D AR     44240
H2 Consumed, Stage 2           0.0476
Heteroatoms (inc H2O)          0.1573                                              PLANT CONSTRUCTION COST ($1000)
C1C3 Gases                     0.0780                                                 LIQUEFACTION                1366197
C4 - 350F                      0.1534          343700       30736                     SOLIDS REMOVAL               140664
350F - 450F Product            0.0775          173642       13609                     HYDROGEN PRODUCTION          681705
450F - 650F Product            0.2359          528545       38557                     BALANCE OF PLANT             464778
650 - 850F Product             0.2162          484402       32951                       TOTAL CONS COST           2653344
   TOTAL C4-850 PRODUCT        0.6830
850F+ Product                  0.0000                                              TOTAL CAPITAL REQUIRED         4286873
Soluble Reject                 0.0822
Unconverted Coal               0.0655                                              OPERATING COSTS,$1000/YR
Ash                            0.1158                                                 COAL COST                    437980
    TOTAL,RAW PRODUCT                         1530288      115853                     OTHER OPERATING COSTS        359917
    TOTAL, HYDROTREATED PRODUCT               1521682      126452                     BY-PRODUCT CREDITS           116421
                                                                                         TOTAL,$1000/YR            681476
  AVERAGE PRODUCT QUALITY
                    CARBON   HYDROGEN    NITROGEN     SULFUR      OXYGEN     API deg      HYDROTREATING COSTS          205401
RAW                  87.21      11.60        0.18       0.08        0.93       24.89        TOTAL INC.HYDROTREAT       886877
HYDR                 87.00      13.00                                          40.00
                                                                                   REQUIRED SELLING PRICE           $/BBL
   DEASHER COST FACTOR         1.00
   COAL CLEANING, %Btu Recovered   100.00 %      0 Flag,1 or 0                        RAW PRODUCT                   33.19
   COAL DRYING FLAG (1 or 0)       0                                                  HYDROTREATED PRODUCT          35.33



                                        SG          C           H          N          S          O         API

                              NAP     0.7660     84.8100     13.8300     0.0500     0.1500     1.1600     53.23
                              MD      0.8740     85.7200     11.9300     0.1400     0.1000     2.1100     30.40
                              HMD     0.9390     87.4500     11.3200     0.1900     0.0600     0.9800     19.19
                              PHD     1.0070     89.1000     10.1900     0.2800     0.0500     0.3000      9.02
```

PROPRIETARY INFORMATION

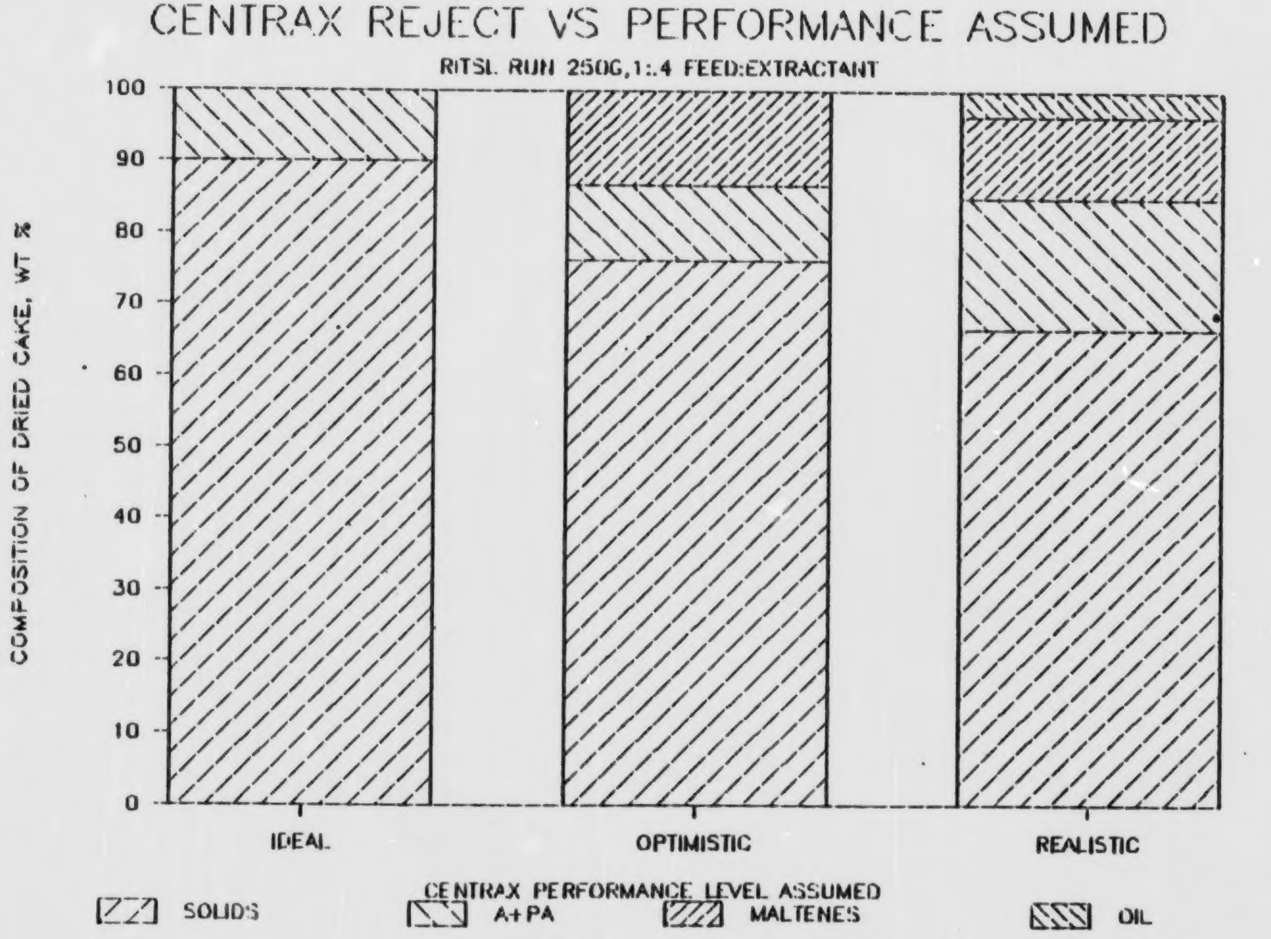

FIGURE 6-8
CENTRAX REJECT VS PERFORMANCE ASSUMED
RTSL RUN 250G, 1:.4 FEED:EXTRACTANT

PROPRIETARY INFORMATION

Figure 6-9 shows the dried cake composition for this case. Now the CENTRAX performance is about the same as was achieved in the other runs. Even with increased extractant, the total number of CENTRAX units required is substantially less than in the previous cases. (i.e., about 2800 units compared to 3800-4400 in earlier cases).

Figures 6-10 and 6-11 show product cost comparisons for the 1:0.4 and 1:0.7 cases. The higher extractant to feed ratio results in slightly lower product costs. Cost savings of 5 percent may be possible under idealized CENTRAX assumptions but for run 250G there is essentially no difference in product costs between CENTRAX and CSD.

Note that improvements in deasher efficiency have a greater impact on raw product costs than on hydrotreated costs. This is because the additional liquid recovered is heavier and required more hydrogen for upgrading.

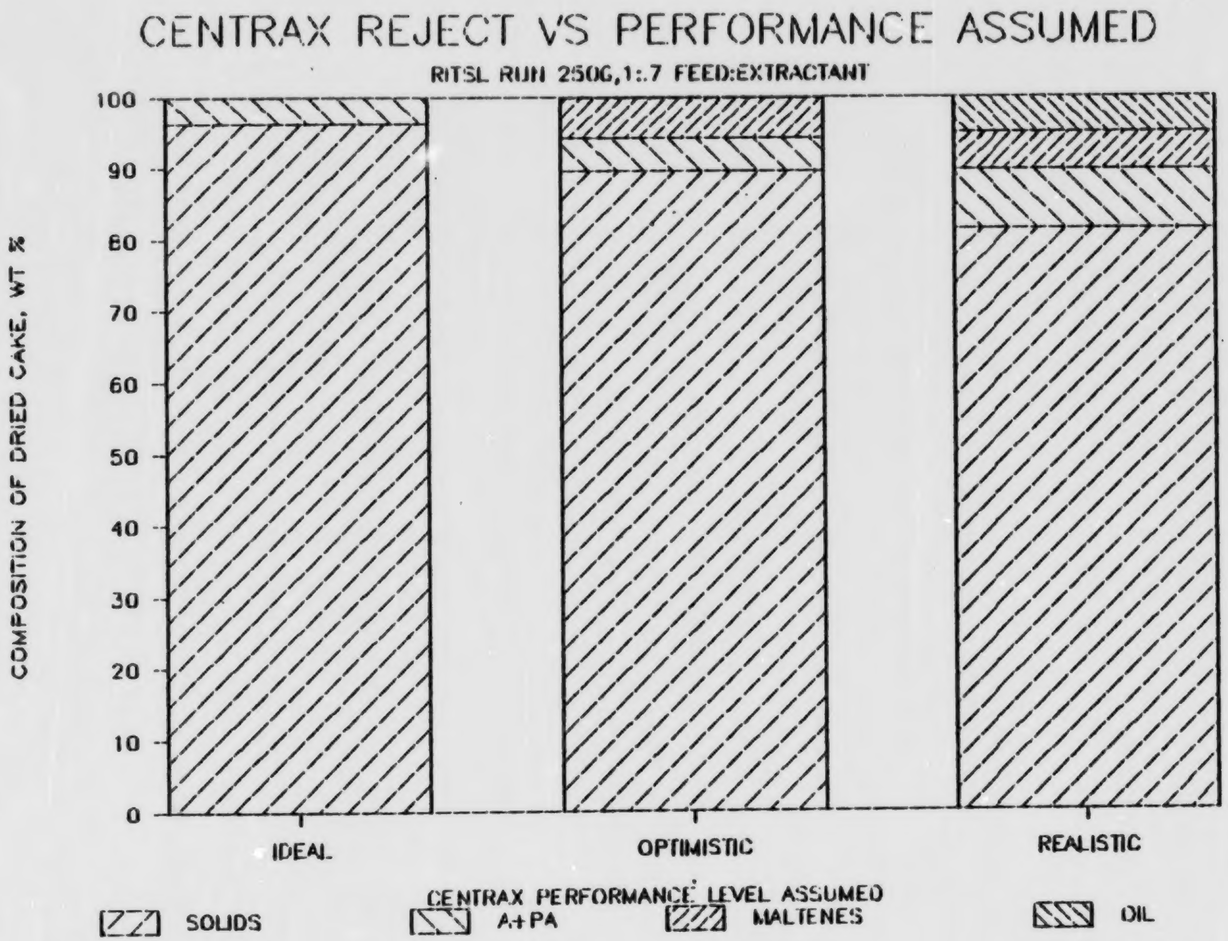

FIGURE 6-9
CENTRAX REJECT VS PERFORMANCE ASSUMED
RITSL RUN 250G, 1:.7 FEED:EXTRACTANT

PROPRIETARY INFORMATION

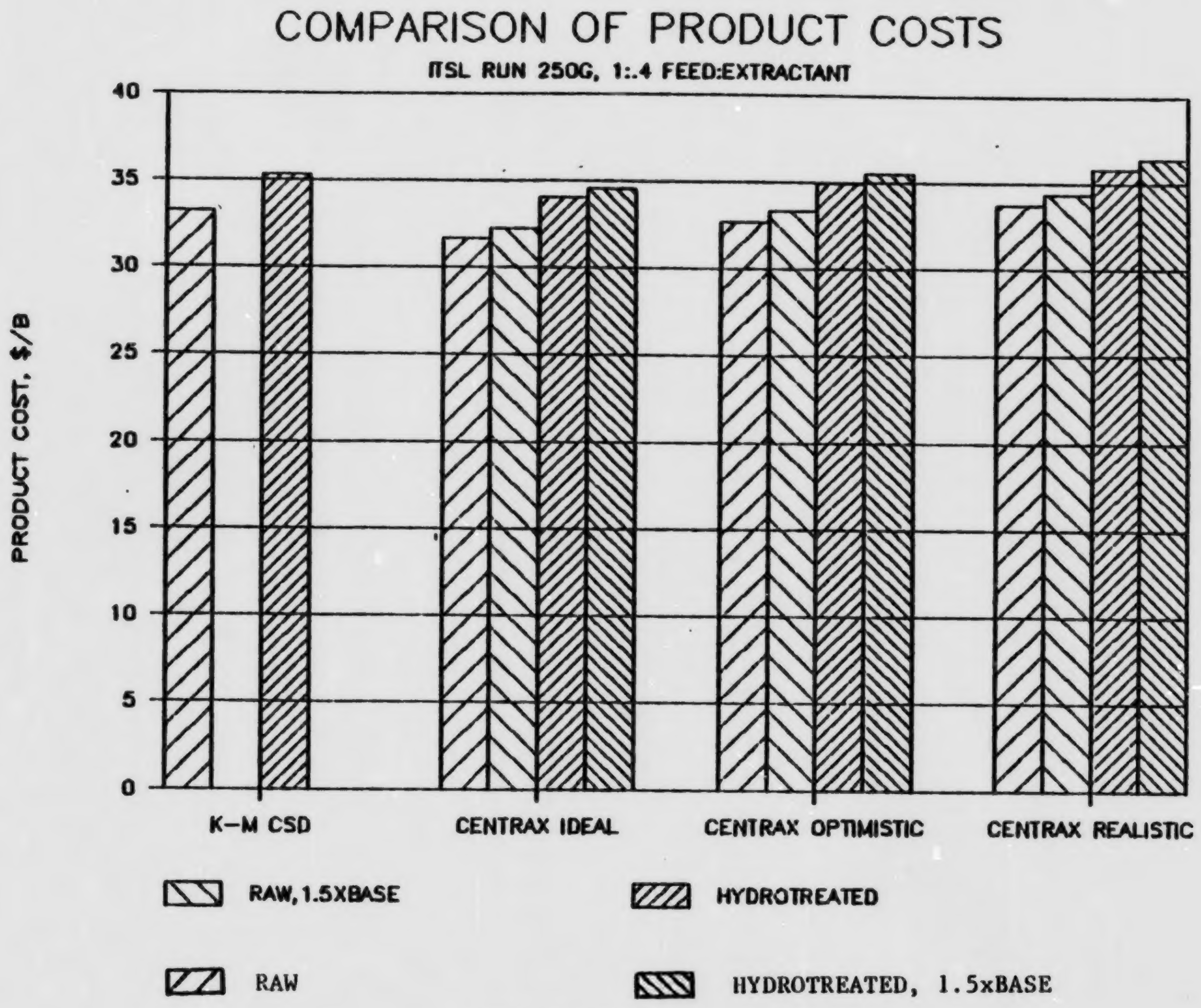

FIGURE 6-10
COMPARISON OF PRODUCT COSTS
ITSL RUN 250G, 1:.4 FEED:EXTRACTANT

PROPRIETARY INFORMATION

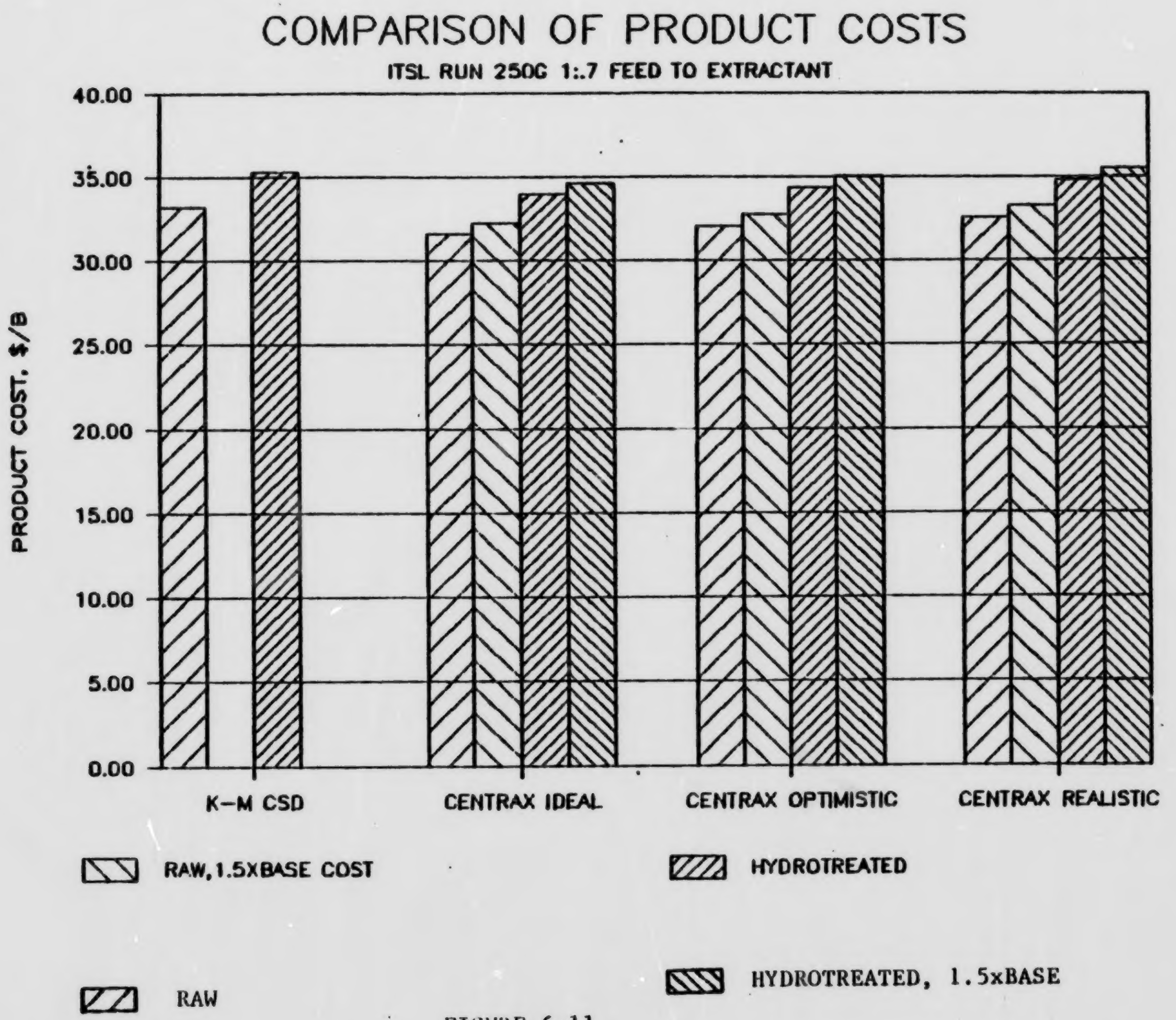

FIGURE 6-11
COMPARISON OF PRODUCT COSTS
ITSL RUN 250G, 1:.7 FEED:EXTRACTANT

PROPRIETARY INFORMATION

7.0 CONCLUSION AND RECOMMENDATIONS

7.1 Conclusion

The MITRE study has been more intensive and far more interesting than we originally envisioned. The development of the MITRE Coal Liquefaction Model and the computer simulation of the multistage CENTRAX system has allowed the process to be evaluated to a much greater depth than was originally planned.

The MITRE analyses presented in Section 6 have indicated that product cost savings in the order of 10 percent compared to CSD deashing could be possible if the ideal level of CENTRAX performance could be achieved. For what we believe to be a more realistic level of CENTRAX performance and cost, savings relative to CSD are small in the ITSL mode, and almost nill in the RITSL or RITSL Ash Recycle mode.

The MITRE assessment of CENTRAX performance has optimisticly assumed that problems of solids cross-over during scraper operation could be solved, and that a 90 percent on line factor for the mechanically complex CENTRAX units can be achieved.

Sandwell makes no particular note of mechanical problems in the pilot unit other than bearing problems which are believed to be solvable. We note that data are reported for only about 8 hours of operation during the .7 month test period. More extensive testing would be expected if the unit were mechanically reliable.

It is notable that in all cases, the CSD performance assumed is the performance actually demonstrated at the relatively large scale

(3 ton/day) Wilsonville pilot plant. CSD performance has been penalized in the MITRE analysis with solvent loss, as well as with the slight increase in UC which occurs in the CSD unit. CSD developers project a substantially improved level of CSD performance once the solvent is tailored to specific process conditions in a commercial facility.

Based on our analysis of CENTRAX, and our knowledge of CSD, we would anticipate that in commercial practice CSD and CENTRAX would both reject a material containing about 75-80% solids in the RITSL configuration. These results differ substantially from the results of the scoping study performed by Sandwell. In their study, the CSD reject was assumed to contain only 57 percent solids compared to the 96 percent assumed for CENTRAX. The CENTRAX advantage was further magnified by the very high ash (20% MAF) lignite assumed. The Sandwell analysis also used the ITSL mode of two stage liquefaction. Our own analysis shows the CENTRAX advantage to be slightly higher in the ITSL mode than in the RITSL or RITSL Ash Recycle Mode. The projected advantage of CENTRAX has thus been slightly eroded by the improvements in two stage processing which have been accomplished since CENTRAX development began.

Run 250G analyzing the RITSL ash recycle mode is of significant interest since this configuration results in lower product cost than the other configurations analyzed. The reduced deasher flow in this configuration substantially reduces the cost of CSD deashing. The CENTRAX unit is afforded a lesser benefit from the reduced deasher flow rate because of its intermittant operation. The number of scraper recycles required is determined by the solids flow, rather than the total deasher flow. Thus, CENTRAX flow times are reduced by the reduced deasher flow, but interrupt time for scraper operation is not. The number of units could be reduced by 30 percent if continuous operation were possible.

Further dramatic reduction in the number of units could be made if operating temperatures could be increased, and larger units typical of scroll type centrifuges were used. Under these conditions, as few as 200-300 operating units might achieve a deashing efficiency comparable to CSD deashing at a substantial cost savings.

Our experience with the study has convinced us that the concept of centrifugal deashing of coal liquids has considerable merit. We question, however, if the semi-continuous CENTRAX unit is a promising implementation of centrifugation, particularly if temperatures must be limited to the 180°C range demonstrated with the pilot unit. At the onset of CENTRAX development, the 180°C feed temperature was believed to be an advantage over conventional scroll centrifuges. In the intervening years, scroll centrifuges operating at twice this temperature have been demonstrated. In this case, as with the change in two stage process flows, the potential advantage of CENTRAX has been overtaken by other technical developments.

7.2 Recommendations

If CENTRAX is to be developed to become a cost competitive deashing system, it will be necessary to demonstrate the ability to achieve performance in the range of the optimistic performance level assumed by MITRE. This will require the demonstration that three stages of centrifugation can be accomplished without significant selective concentration of A and PA material in the underflow, and that a high recovery of oil in the compacted cake can be achieved.

Problems with solids cross-over during scraper operation must also be solved. We are not optimistic that these problems will yield to further development and this effort is thus not recommended. The problem of liquid recovery is of particular interest. High recoveries will necessarily require coking conditions. In the RITSL case, if coking is envisioned, the deasher feed could be coked without prior centrifugation. This option may in fact be very attractive in the RITSL Ash Recycle mode. (Run 250G)

We recommend that further development of centrifugal systems for deashing coal derived and coprocessing liquids concentrate on the demonstration of high temperature scroll-type centrifuges. These units are believed to have considerable potential.

APPENDIX A

```
                       TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-   244B            70.6000 SOLID REJECT
                               ILLINOIS # 6 RUN 15244B, K-M CRITICAL SOLVENT DEASHING

                                                          COMMERCIAL PLANT OUTPUT
  YIELD STRUCTURE,              WT% MAF COAL               LBS/HR      BBLS/DAY          COAL TO LIQUEFACTION, T/D AR     33515
                                                                                         COAL TO STEAM,        T/D AR      1996
  H2 Consumed, Stage 1             0.0122                                                COAL TO HYDROGEN,     T/D AR      3522
  H2 Consumed, Stage 2             0.0389                                                    TOTAL COAL        T/D AR     39033
  Heteroatoms (inc H2O)            0.1397
  C1C3 Gases                       0.0526                                                PLANT CONSTRUCTION COST ($1000)
  C4 - 350F                        0.0965                                                    LIQUEFACTION                1476595
  350F - 450F Product              0.0730              215942       10253                    SOLIDS REMOVAL               278959
  450F - 650F Product              0.2103              163355       11870                    HYDROGEN PRODUCTION          578755
  650 - 850F Product               0.2026              470596       34195                    BALANCE OF PLANT             443473
     TOTAL C4-850F PRODUCT         0.6624              632449       42494                      TOTAL CONS COST           2777782
  850F+ Product                    0.0000
  Soluble Reject                   0.0922                                                TOTAL CAPITAL REQUIRED          4457159
  Unconverted Coal                 0.1042
  Ash                              0.1172                                                OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                                                                       COAL COST                    386424
     TOTAL, HYDROTREATED PRODUCT                      1482341      106812                    OTHER OPERATING COSTS        385811
                                                      1476232      122675                    BY-PRODUCT CREDITS            39592
                                                                                               TOTAL,$1000/YR             732642
  AVERAGE PRODUCT QUALITY
                    CARBON    HYDROGEN     NITROGEN     SULFUR      OXYGEN      API deg          HYDROTREATING COSTS      329748
RAW                  87.34     10.60         0.24        0.21        1.60       17.35              TOTAL INC.HYDROTREAT   1062391
HYDR                 87.00     13.00                                            40.00
                                                                                         REQUIRED SELLING PRICE           $/BBL
  DEASHER COST FACTOR             1.00
  COAL CLEANING, %Btu Recovered  100.00 %        0 Flag,1 or 0                               RAW PRODUCT                   38.11
  COAL DRYING FLAG (1 or 0)          0                                                       HYDROTREATED PRODUCT          48.33


                            SG          C .          H          N          S          O        API

              NAP         0.8104     85.2100     12.8600     0.0845     0.3600     1.5000     43.11
              MD          0.9427     86.3400     10.7300     0.2300     0.2200     2.4800     18.60
              HHD         0.9427     86.3400     10.7300     0.2300     0.2200     2.4800     18.60
              PWD         1.0195     89.0700      9.6900     0.3100     0.1600     0.7600      7.29
```

```
TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    244B          96.4100 SOLID REJECT
          ILLINOIS # 6 RUN 15244B CENTRAX DEASHING

YIELD STRUCTURE,            #/# MAF COAL        COMMERCIAL PLANT OUTPUT        COAL TO LIQUEFACTION, T/D AR    33515
                                                 LBS/HR     BBLS/DAY          COAL TO STEAM,        T/D AR     2172
 H2 Consumed, Stage 1          0.0129                                          COAL TO HYDROGEN,     T/D AR     9149
 H2 Consumed, Stage 2          0.0443                                             TOTAL COAL         T/D AR    44836
 Heteroatoms (inc H2O)         0.1482
 C1C3 Gases                    0.0577                                         PLANT CONSTRUCTION COST ($1000)
 C4 - 350F                     0.1019            220067      19278               LIQUEFACTION                1553045
 350F - 450F Product           0.0771            172527      12536               SOLIDS REMOVAL               320331
 450F - 650F Product           0.2221            497021      36115               HYDROGEN PRODUCTION          612855
 650 - 850F Product            0.3095            871614      58564               BALANCE OF PLANT             469147
    TOTAL C4-850F PRODUCT      0.7906                                               TOTAL CONS COST          2955377
 850F+ Product                 0.0000
 Soluble Reject                0.0064                                         TOTAL CAPITAL REQUIRED          4743844
 Unconverted Coal              0.0539
 Ash                           0.1172                                         OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                           1769229     126493              COAL COST                     443872
    TOTAL, HYDROTREATED PRODUCT                 1765954     146751              OTHER OPERATING COSTS         388748
                                                                                BY-PRODUCT CREDITS             63566
AVERAGE PRODUCT QUALITY                                                             TOTAL,$1000/YR            769054
                 CARBON   HYDROGEN   NITROGEN   SULFUR   OXYGEN   API deg
RAW               87.54    10.49      0.25       0.21     1.51     16.19        HYDROTREATING COSTS           411063
HYDR              87.00    13.00                                   40.00           TOTAL INC.HYDROTREAT      1180118

 DEASHER COST FACTOR            1.00                                          REQUIRED SELLING PRICE           $/BBL
 COAL CLEANING, %Btu Recovered  100.00 %      0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)      0                                               RAW PRODUCT                    33.99
                                                                                HYDROTREATED PRODUCT           37.79


                      SG         C          H          N         S         O        API

          NAP       0.8104    85.2100    12.8600    0.0845    0.3600    1.5000    43.11
          MD        0.9427    86.3400    10.7300    0.2300    0.2200    2.4800    18.60
          HHD       0.9427    86.3400    10.7300    0.2300    0.2200    2.4800    18.60
          PHD       1.0195    89.0700     9.6900    0.3100    0.1600    0.7600     7.29
          TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    244B           96.4100 SOLID REJECT
                      ILLINOIS # 6 RUN TS244B CRITICAL DEASHING

 YIELD STRUCTURE,          W/W MAF COAL          COMMERCIAL PLANT OUTPUT         COAL TO LIQUEFACTION, T/D AR     33515
                                                  LBS/HR      BBLS/DAY           COAL TO STEAM,        T/D AR      2172
 H2 Consumed, Stage 1         0.0129                                             COAL TO HYDROGEN,     T/D AR      9149
 H2 Consumed, Stage 2         0.0443                                               TOTAL COAL          T/D AR     44836
 Heteroatoms (inc H2O)        0.1482
 C1C3 Gases                   0.0577                                             PLANT CONSTRUCTION COST ($1000)
 C4 - 350F                    0.1019                                               LIQUEFACTION                 1553045
 350F - 450F Product          0.0771             228067        19278               SOLIDS REMOVAL                480496
 450F - 650F Product          0.2221             172527        12536               HYDROGEN PRODUCTION           612855
 650 - 850F Product           0.3895             497021        36115               BALANCE OF PLANT              469147
   TOTAL C4-850F PRODUCT      0.7906             871614        58564                 TOTAL CONS COST            3115543
 850F+ Product                0.0000
 Soluble Reject               0.0064                                             TOTAL CAPITAL REQUIRED         4982841
 Unconverted Coal             0.0539
 Ash                          0.1172                                             OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                           1769229       126493               COAL COST                     443872
    TOTAL, HYDROTREATED PRODUCT                 1765954       146751               OTHER OPERATING COSTS         399759
                                                                                   BY-PRODUCT CREDITS             63566
 AVERAGE PRODUCT QUALITY                                                             TOTAL,$1000/YR              780066
                 CARBON   HYDROGEN    NITROGEN    SULFUR      OXYGEN    API deg
 RAW              87.54     10.49                                                  HYDROTREATING COSTS           411063
 HYDR             87.00     13.00        0.25       0.21        1.51     16.19       TOTAL INC.HYDROTREAT       1191129
                                                                         40.00
   DEASHER COST FACTOR        1.50                                               REQUIRED SELLING PRICE           $/BBL
   COAL CLEANING, %Btu Recovered 100.00 %     0 Flag,1 or 0
   COAL DRYING FLAG (1 or 0)     0                                                 RAW PRODUCT                    35.04
                                                                                   HYDROTREATED PRODUCT           38.69

                              SG        C          H         N         S         O        API

              NAP          0.8104    85.2100    12.8600    0.0045    0.3600    1.5000    43.11
              MD           0.9427    86.3400    10.7300    0.2300    0.2200    2.4800    18.60
              MHD          0.9427    86.3400    10.7300    0.2300    0.2200    2.4800    18.60
              PHD          1.0195    89.0700     9.6900    0.3100    0.1600    0.7600     7.29
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                    TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-   244B            30.1700 SOLID REJECT
                                    ILLINOIS # 6 RUN TX244B CENTRAL DEASHING

 YIELD STRUCTURE,                 WTS MAF COAL          COMMERCIAL PLANT OUTPUT          COAL TO LIQUEFACTION, T/D AR     33515
                                                          LBS/HR     BBLS/DAY            COAL TO STEAM,        T/D AR      2772
 H2 Consumed, Stage 1               0.0124                                               COAL TO HYDROGEN,     T/D AR      6314
 H2 Consumed, Stage 2               0.0420                                                  TOTAL COAL         T/D AR     42601
 Heteroatoms (inc H2O)              0.1125
 C1C3 Gases                         0.0551                                               PLANT CONSTRUCTION COST ($1000)
 C4 - 350F Product                  0.0981                                                  LIQUEFACTION                1533428
 350F - 450F Product                0.0742               219413      18546                  SOLIDS REMOVAL               339686
 450F - 650F Product                0.2137               165381      12061                  HYDROGEN PRODUCTION          555296
 650 - 850F Product                 0.3506               478161      34745                  BALANCE OF PLANT             463895
    TOTAL C4-850F PRODUCT           0.7445               802395      63913                    TOTAL CONS COST           2892305
 850F+ Product                      0.0000
 Soluble Reject                     0.0226                                               TOTAL CAPITAL REQUIRED          4599091
 Unconverted Coal                   0.0838
 Ash                                0.1172                                               OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                                                                       COAL COST                    421747
    TOTAL, HYDROTREATED PRODUCT                         1665949     119265                  OTHER OPERATING COSTS        387316
                                                        1662235     130132                  BY-PRODUCT CREDITS            52568
 AVERAGE PRODUCT QUALITY                                                                      TOTAL,$1000/YR             756495
                     CARBON   HYDROGEN     NITROGEN    SULFUR      OXYGEN     API deg
RAW                   87.51     10.51          0.25      0.21        1.52       16.38        HYDROTREATING COSTS          384319
HYDR                  87.00     13.00                                           40.00          TOTAL INC.HYDROTREAT       1140814

 DEASHER COST FACTOR                1.00                                                 REQUIRED SELLING PRICE          $/BBL
 COAL CLEANING, MBtu Recovered    100.00 %        0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)             0                                                    RAW PRODUCT                   35.58
                                                                                            HYDROTREATED PRODUCT          39.15


                           SG           C            H           N          S          O         API

           NAP          0.8104      85.2100      12.8600     0.0846     0.3600     1.5000      43.11
           MD           0.9427      86.3100      10.7300     0.2300     0.2200     2.4000      18.60
           HND          0.9427      86.3100      10.7300     0.2300     0.2200     2.4000      18.60
           PHD          1.0136      89.0100       9.6300     0.3100     0.1600     0.7600       7.28
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          TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    2440           90.1700 SOLID REJECT
                         ILLINOIS # 6 RUN 132440 CENTRAX DEASHING

 YIELD STRUCTURE,             WT% MAF COAL            COMMERCIAL PLANT OUTPUT        COAL TO LIQUEFACTION, T/D AR     33515
                                                       LBS/HR      BBLS/DAY        . COAL TO STEAM,        T/D AR      2172
 H2 Consumed, Stage 1           0.0124                                               COAL TO HYDROGEN,     T/D AR      6914
 H2 Consumed, Stage 2           0.0420                                                  TOTAL COAL         T/D AR     42601
 Heteroatoms (inc H2O)          0.1425
 C1C3 Gases                     0.0551                                              PLANT CONSTRUCTION COST ($1000)
 C4 - 350F Product              0.0981                                                 LIQUEFACTION                 1533428
 350F - 450F Product            0.0742                 219413       18546              SOLIDS REMOVAL                509528
 450F - 650F Product            0.2137                 165981       12061              HYDROGEN PRODUCTION           595296
 650 - 850F Product             0.3586                 478161       34745              BALANCE OF PLANT              463595
    TOTAL C4-850F PRODUCT       0.7445                 802395       53913                TOTAL CONS COST            3101848
 850F+ Product                  0.0000
                                                                                    TOTAL CAPITAL REQUIRED          4952528

 Ash                            0.1172
    TOTAL,RAW PRODUCT                                                                  COAL COST                     421747
    TOTAL, HYDROTREATED PRODUCT                       1665949      119265              OTHER OPERATING COSTS         398993
                                                      1662235      138132              BY-PRODUCT CREDITS             52569
 AVERAGE PRODUCT QUALITY                                                                 TOTAL,$1000/YR              768171
                      CARBON    HYDROGEN     NITROGEN    SULFUR     OXYGEN    API deg
RAW                    87.51     10.51         0.25       0.21       1.52     16.38     HYDROTREATING COSTS           384319
HYDR                   87.00     13.00                                         40.00       TOTAL INC.HYDROTREAT     1152490

  DEASHER COST FACTOR           1.50                                                REQUIRED SELLING PRICE          $/BBL
  COAL CLEANING, %Btu Recovered 100.00 %        0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)       0                                                    RAW PRODUCT                    36.76
                                                                                       HYDROTREATED PRODUCT           40.17


                            SG           C            H          N         S         O          API

           NAP            0.8104      85.2100      12.8600    0.0845    0.3600    1.5000      43.11
           RD             0.9427      86.3400      10.7300    0.2300    0.2200    2.4000      18.60
           MID            0.9427      86.3400      10.7300    0.2300    0.2200    2.4000      18.60
           FUO            1.0195      89.0700       9.6900    0.3100    0.1600    0.7600       7.29
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                     TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    244B          81.8500 SOLID REJECT
                               ILLINOIS # 6 RUN 15244B CENTRAX DEASHING

  YIELD STRUCTURE,            WT% MAF COAL            COMMERCIAL PLANT OUTPUT         COAL TO LIQUEFACTION, T/D AR     33515
                                                       LBS/HR      BBLS/DAY           COAL TO STEAM,        T/D AR      2172
  H2 Consumed, Stage 1           0.0124                                               COAL TO HYDROGEN,     T/D AR      5992
  H2 Consumed, Stage 2           0.0412                                                  TOTAL COAL         T/D AR     41679
  Heteroatoms (inc H2O)          0.1423
  C1C3 Gases                     0.0546                                               PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product              0.0981                                                  LIQUEFACTION                 1515491
  350F - 450F Product            0.0742                219413       18546                SOLIDS REMOVAL                339686
  450F - 650F Product            0.2137                165981       12061                HYDROGEN PRODUCTION           590673
  650 - 850F Product             0.3351                478161       34745                BALANCE OF PLANT              460460
     TOTAL C4-850F PRODUCT       0.7211                749968       50390                   TOTAL CONS COST           2906309
  850F+ Product                  0.0000
  Soluble Reject                 0.0459                                               TOTAL CAPITAL REQUIRED          4656557
  Unconverted Coal               0.0898
  Ash                            0.1172                                               OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                                1613522      115742                COAL COST                     412620
     TOTAL, HYDROTREATED PRODUCT                      1600990      133707                OTHER OPERATING COSTS         385362
                                                                                         BY-PRODUCT CREDITS             49469
  AVERAGE PRODUCT QUALITY                                                                   TOTAL,$1000/YR             748514
                      CARBON   HYDROGEN     NITROGEN    SULFUR     OXYGEN     API deg
RAW                    87.46     10.54          0.25      0.21       1.55       16.68    HYDROTREATING COSTS           368153
HYDR                   87.00     13.00                                          40.00       TOTAL INC.HYDROTREAT      1116666

  DEASHER COST FACTOR               1.00                                              REQUIRED SELLING PRICE           $/BBL
  COAL CLEANING, %Btu Recovered   100.00 %        0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)            0                                                 RAW PRODUCT                    36.30
                                                                                         HYDROTREATED PRODUCT           39.77


                           SG           C            H          N          S          O         API

              NAP        0.8104      85.2100      12.8600     0.0845     0.3600     1.5000      43.11
              MD         0.9427      86.3400      10.7300     0.2300     0.2200     2.4800      18.60
              HHD        0.9427      86.3400      10.7300     0.2300     0.2200     2.4800      18.60
              PHD        1.0195      89.0700       9.6900     0.3100     0.1600     0.7600       7.29
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                    TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    244B          81.8500 SOLID REJECT
                              ILLINOIS # 6 RUN 15244B CENTRAX DEASHING

                                                    COMMERCIAL PLANT OUTPUT
   YIELD STRUCTURE,             WTS MAF COAL                                           COAL TO LIQUEFACTION, T/D AR     33515
                                                    LBS/HR      BBLS/DAY               COAL TO STEAM,        T/D AR      2172
   H2 Consumed, Stage 1           0.0124                                               COAL TO HYDROGEN,     T/D AR      5992
   H2 Consumed, Stage 2           0.0412                                                  TOTAL COAL         T/D AR     41679
   Heteroatoms (inc H2O)          0.1423
   C1C3 Gases                     0.0546                                               PLANT CONSTRUCTION COST ($1000)
   C4 - 350F Product              0.0981            219413      18546                     LIQUEFACTION                1515491
   350F - 450F Product            0.0742            165981      12061                     SOLIDS REMOVAL               509528
   450F - 650F Product            0.2137            478161      34745                     HYDROGEN PRODUCTION          590673
   650 - 850F Product             0.3351            749968      50390                     BALANCE OF PLANT             460460
      TOTAL C4-850F PRODUCT       0.7211                                                    TOTAL CONS COST           3076152
   850F+ Product                  0.0000
   Soluble Reject                 0.0459                                               TOTAL CAPITAL REQUIRED         4909994
   Unconverted Coal               0.0898
   Ash                            0.1172                                               OPERATING COSTS,$1000/YR
      TOTAL,RAW PRODUCT                                                                   COAL COST                    412620
      TOTAL, HYDROTREATED PRODUCT                   1613522     115742                    OTHER OPERATING COSTS        397039
                                                    1608990     133707                    BY-PRODUCT CREDITS            49469
  AVERAGE PRODUCT QUALITY                                                                    TOTAL,$1000/YR            760190
                       CARBON   HYDROGEN      NITROGEN    SULFUR      OXYGEN     API deg
RAW                     87.46     10.54         0.25        0.21        1.55       16.60      HYDROTREATING COSTS          368153
HYDR                    87.00     13.00                                            40.00        TOTAL INC.HYDROTREAT      1128343

   DEASHER COST FACTOR            1.50                                                 REQUIRED SELLING PRICE          $/BBL
   COAL CLEANING, %Btu Recovered  100.00 %      0 Flag,1 or 0
   COAL DRYING FLAG (1 or 0)         0                                                    RAW PRODUCT                   37.51
                                                                                          HYDROTREATED PRODUCT          40.82


                          SG          C            H          N          S          O         API

           NAP          0.8104     85.2100      12.8600     0.0045     0.3600     1.5000     43.11
           MD           0.9427     86.3400      10.7300     0.2300     0.2200     2.4800     18.60
           HMD          0.9427     86.3400      10.7300     0.2300     0.2200     2.4800     18.60
           PHD          1.0195     89.0700       9.6900     0.3100     0.1600     0.7600      7.29
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                    TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250D            62.8000 SOLID REJECT
                              ILLINOIS # 6 RUN TS250D,K-M CRITICAL SOLVENT DEASHING

                                                     COMMERCIAL PLANT OUTPUT
 YIELD STRUCTURE,                   WT% MAF COAL      LBS/HR      BBLS/DAY              COAL TO LIQUEFACTION, T/D AR     33135
                                                                                        COAL TO STEAM,        T/D AR      2418
 H2 Consumed, Stage 1                 0.0166                                            COAL TO HYDROGEN,     T/D AR      4592
 H2 Consumed, Stage 2                 0.0439                                               TOTAL COAL         T/D AR     40145
 Heteroatoms (inc H2O)                0.1475
 C1C3 Gases                           0.0695                                            PLANT CONSTRUCTION COST ($1000)
 C4 - 350F                            0.1330                                               LIQUEFACTION                1301028
 350F - 450F Product                  0.0708       294248        25841                     SOLIDS REMOVAL               220873
 450F - 650F Product                  0.2130       156637        12138                     HYDROGEN PRODUCTION          643616
 650 - 850F Product                   0.2087       471239        34561                     BALANCE OF PLANT             447446
    TOTAL C4-850F PRODUCT             0.6255       461636        31559                       TOTAL CONS COST           2612964
 850F+ Product                        0.0000
 Soluable Reject                      0.1295                                            TOTAL CAPITAL REQUIRED         4207998
 Undesolved Coal                      0.0886
                                                                                        OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                             1383759       104098                     COAL COST                    397434
    TOTAL, HYDROTREATED PRODUCT                   1379577       114643                     OTHER OPERATING COSTS        356308
                                                                                           BY-PRODUCT CREDITS            85092
 AVERAGE PRODUCT QUALITY                                                                     TOTAL,$1000/YR             668650
                          CARBON   HYDROGEN     NITROGEN     SULFUR      OXYGEN     API deg
AW                         87.44     11.79        0.23        0.10        0.44       23.90     HYDROTREATING COSTS          163036
'DR                        87.00     13.00                                           40.00       TOTAL INC.HYDROTREAT       831686

 DEASHER COST FACTOR                    1.00                                                   REQUIRED SELLING PRICE      $/BBL
 COAL CLEANING, %Btu Recovered        100.00 %          0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)                 0                                                      RAW PRODUCT               36.25
                                                                                                  HYDROTREATED PRODUCT      37.22


                                    SG          C            H          N          S          O          API

                      NAP         0.7800      85.4100      14.2600    0.0400     0.1800     0.1100      49.91
                      ND          0.8840      86.1900      12.4400    0.1300     0.1300     1.1100      28.57
                      MID         0.9340      87.8100      11.5300    0.2500     0.0800     0.3300      20.00
                      PRD         1.0020      88.7700      10.2700    0.3600     0.0700     0.5300       9.72
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             TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    2500           95.2400 SOLID REJECT
                         ILLINOIS # 6 RUN 152500 CENTRAX DEASHING, IDEAL CASE

 YIELD STRUCTURE,              #/# MAF COAL          COMMERCIAL PLANT OUTPUT         COAL TO LIQUEFACTION, T/D AR      33135
                                                      LBS/HR      BBLS/DAY            COAL TO STEAM,        T/D AR       2625
  H2 Consumed, Stage 1           0.0174                                               COAL TO HYDROGEN,     T/D AR      11236
  H2 Consumed, Stage 2           0.0504                                                  TOTAL COAL         T/D AR      46997
  Heteroatoms (inc H2O)          0.1552
  C1C3 Gases                     0.0757                                               PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product              0.1391                                                  LIQUEFACTION                 1373672
  350F - 450F Product            0.0741              307808       27032                  SOLIDS REMOVAL                271480
  450F - 650F Product            0.2228              163855       12697                  HYDROGEN PRODUCTION           682070
  650 - 850F Product             0.3450              492955       36154                  BALANCE OF PLANT              477333
     TOTAL C4-850F PRODUCT       0.7810              763281       52180                    TOTAL CONS COST            2804555
  850F+ Product                  0.0000
  Soluble Reject                 0.0088                                               TOTAL CAPITAL REQUIRED          4528067
  Unconverted Coal               0.0466
  Ash                            0.1300                                               OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                               1727899      128063                 COAL COST                     465272
     TOTAL, HYDROTREATED PRODUCT                     1726940      143509                 OTHER OPERATING COSTS         377597
                                                                                         BY-PRODUCT CREDITS            113603
  AVERAGE PRODUCT QUALITY                                                                   TOTAL,$1000/YR             729266
                   CARBON   HYDROGEN     NITROGEN    SULFUR     OXYGEN     API deg
RAW                 87.65     11.55        0.25       0.10       0.45       21.60         HYDROTREATING COSTS          242677
HYDR                87.00     13.00                                         40.00           TOTAL INC.HYDROTREAT       971942

  DEASHER COST FACTOR            1.00                                                 REQUIRED SELLING PRICE          $/BBL
  COAL CLEANING, %Btu Recovered  100.00 %          0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)          0                                                   RAW PRODUCT                  31.94
                                                                                         HYDROTREATED PRODUCT         33.62

                                     SG          C           H           N          S          O         API

                    NAP            0.7800     85.4100     14.2600     0.0400     0.1800     0.1100     49.91
                    JD             0.8840     86.1900     12.4400     0.1300     0.1300     1.1100     28.57
                    MID            0.9340     87.8100     11.5300     0.2500     0.0800     0.3300     20.00
                    FUO            1.0020     88.7700     10.2700     0.3600     0.0700     0.5300      9.72
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            TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    2500          95.2400 SOLID REJECT
                        ILLINOIS # 6 RUN IS2500 CENTRAX DEASHING, IDEAL CASE

 YIELD STRUCTURE,              B/B MAF COAL           COMMERCIAL PLANT OUTPUT        COAL TO LIQUEFACTION, T/D AR    33135
                                                        LBS/HR     BBLS/DAY           COAL TO STEAM,        T/D AR     2625
 H2 Consumed, Stage 1             0.0174                                              COAL TO HYDROGEN,     T/D AR    11236
 H2 Consumed, Stage 2             0.0504                                                 TOTAL COAL         T/D AR    46996
 Heteroatoms (inc H2O)            0.1552
 C1C3 Gases                       0.0757                                              PLANT CONSTRUCTION COST ($1000)
 C4 - 350F Product                0.1391                                                 LIQUEFACTION                1373672
 350F - 450F Product              0.0741              307808      27032                  SOLIDS REMOVAL               407220
 450F - 650F Product              0.2228              163855      12697                  HYDROGEN PRODUCTION          682070
 650 - 850F Product               0.3450              492955      36154                  BALANCE OF PLANT             477333
    TOTAL C4-850F PRODUCT         0.7810              763281      52180                    TOTAL CONS COST           2940295
 850F+ Product                    0.0000
 Soluble Reject                   0.0088                                              TOTAL CAPITAL REQUIRED         4730616
 Unconverted Coal                 0.0466
 Ash                              0.1300                                              OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                                                                    COAL COST                    465272
    TOTAL, HYDROTREATED PRODUCT                      1727899     128063                  OTHER OPERATING COSTS        386929
                                                     1726940     143509                  BY-PRODUCT CREDITS           113603
 AVERAGE PRODUCT QUALITY                                                                   TOTAL,$1000/YR             738598
                    CARBON   HYDROGEN    NITROGEN     SULFUR     OXYGEN     API deg
RAW                  87.65     11.55        0.25        0.10       0.45       21.60       HYDROTREATING COSTS          242677
HYDR                 87.00     13.00                                          40.00         TOTAL INC.HYDROTREAT       981275

 DEASHER COST FACTOR               1.50                                              REQUIRED SELLING PRICE          $/BBL
 COAL CLEANING, %Btu Recovered   100.00 %           0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)            0                                                 RAW PRODUCT                    32.81
                                                                                        HYDROTREATED PRODUCT           34.41


                          SG           C            H          N          S          O         API

              NAP       0.7800      85.4100      14.2600     0.0400     0.1800     0.1100     49.91
              MD        0.8840      86.1900      12.4400     0.1300     0.1300     1.1100     28.57
              HHD       0.9340      87.8100      11.5300     0.2500     0.0800     0.3300     20.00
              PHD       1.0020      88.7700      10.2700     0.3600     0.0700     0.5300      9.72
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TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-   250D          87.0300 SOLID REJECT
               ILLINOIS # 6 RUN TS250D CENTRAX DEASHING, OPTIMISTIC CASE

                                          COMMERCIAL PLANT OUTPUT            COAL TO LIQUEFACTION, T/D AR    33135
YIELD STRUCTURE,          WT% MAF COAL     LBS/HR     BBLS/DAY               COAL TO STEAM,        T/D AR     2625
                                                                             COAL TO HYDROGEN,     T/D AR     8933
H2 Consumed, Stage 1         0.0168                                             TOTAL COAL         T/D AR    44694
H2 Consumed, Stage 2         0.0479
Heteroatoms (inc H2O)        0.1500                                          PLANT CONSTRUCTION COST ($1000)
C1C3 Gases                   0.0727                                             LIQUEFACTION                1357256
C4 - 350F Product            0.1346        297767       26150                   SOLIDS REMOVAL               287990
350F - 450F Product          0.0716        150510       12283                   HYDROGEN PRODUCTION          663378
450F - 650F Product          0.2155        476875       34974                   BALANCE OF PLANT             471934
650 - 850F Product           0.3102        686371       46923                     TOTAL CONS COST           2780559
   TOTAL C4-850F PRODUCT     0.7320
850F+ Product                0.0000                                          TOTAL CAPITAL REQUIRED         4482024
Soluble Reject               0.0310
Unconverted Coal             0.0777                                          OPERATING COSTS,$1000/YR
Ash                          0.1300                                             COAL COST                    442469
    TOTAL,RAW PRODUCT                     1619523      120330                   OTHER OPERATING COSTS        376004
    TOTAL, HYDROTREATED PRODUCT           1617961      134453                   BY-PRODUCT CREDITS           101331
                                                                                  TOTAL,$1000/YR             717142
 AVERAGE PRODUCT QUALITY
                      CARBON  HYDROGEN    NITROGEN    SULFUR     OXYGEN    API deg     HYDROTREATING COSTS       221380
RAW                    87.62     11.59        0.25      0.10       0.45      21.98        TOTAL INC.HYDROTREAT     938522
HYDR                   87.00     13.00                                       40.00
                                                                             REQUIRED SELLING PRICE           $/BBL
  DEASHER COST FACTOR            1.00
  COAL CLEANING, %Btu Recovered  100.00 %      0 Flag,1 or 0                    RAW PRODUCT                   33.52
  COAL DRYING FLAG (1 or 0)          0                                          HYDROTREATED PRODUCT          34.99


                       SG          C           H          N          S          O         API

          NAP        0.7800     85.4100     14.2600     0.0400     0.1800     0.1100     49.91
          MD         0.8840     86.1900     12.4400     0.1300     0.1300     1.1100     28.57
          HHD        0.9340     87.8100     11.5300     0.2500     0.0800     0.3300     20.00
          PHD        1.0020     88.7700     10.2700     0.3600     0.0700     0.5300      9.72
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                    TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250D          87.0300 SOLID REJECT
                              ILLINOIS # 6 RUN IS250D CENTRAX DEASHING, OPTIMISTIC CASE

  YIELD STRUCTURE,             WT% MAF COAL          COMMERCIAL PLANT OUTPUT              COAL TO LIQUEFACTION, T/D AR     33135
                                                      LBS/HR      BBLS/DAY                 COAL TO STEAM,        T/D AR      2625
  H2 Consumed, Stage 1             0.0168                                                  COAL TO HYDROGEN,     T/D AR      8933
  H2 Consumed, Stage 2             0.0479                                                      TOTAL COAL        T/D AR     44694
  Heteroatoms (inc H2O)            0.1500
  C1C3 Gases                       0.0727                                                 PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product                0.1346            297767       26150                      LIQUEFACTION                  1357256
  350F - 450F Product              0.0716            158510       12283                      SOLIDS REMOVAL                 431985
  450F - 650F Product              0.2155            476875       34974                      HYDROGEN PRODUCTION            663378
  650 - 850F Product               0.3102            686371       46923                      BALANCE OF PLANT               471934
     TOTAL C4-850F PRODUCT         0.7320                                                      TOTAL CONS COST             2924554
  850F+ Product                    0.0000
  Soluble Reject                   0.0310                                                 TOTAL CAPITAL REQUIRED           4696892
  Unconverted Coal                 0.0777
  Ash                              0.1300                                                 OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                               1619523      120330                     COAL COST                      442469
     TOTAL, HYDROTREATED PRODUCT                     1617961      134453                     OTHER OPERATING COSTS          385904
                                                                                             BY-PRODUCT CREDITS             101331
 AVERAGE PRODUCT QUALITY                                                                        TOTAL,$1000/YR              727042
                   CARBON   HYDROGEN     NITROGEN     SULFUR      OXYGEN     API deg
RAW                 87.62     11.59         0.25        0.10        0.45       21.98          HYDROTREATING COSTS            221380
HYDR                87.00     13.00                                            40.00            TOTAL INC.HYDROTREAT         948422

  DEASHER COST FACTOR              1.50                                                   REQUIRED SELLING PRICE             $/BBL
  COAL CLEANING, %Btu Recovered  100.00 %        0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)          0                                                       RAW PRODUCT                     34.51
                                                                                             HYDROTREATED PRODUCT            35.88


                               SG           C              H           N          S          O          API

                 NAP         0.7800      85.4100        14.2600      0.0400     0.1800     0.1100       49.91
                 HD          0.8840      86.1900        12.4400      0.1300     0.1300     1.1100       28.57
                 MID         0.9340      87.8100        11.5300      0.2500     0.0800     0.3300       20.00
                 PID         1.0020      88.7700        10.2700      0.3600     0.0700     0.5300        9.72
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TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-   2500          70.5300 SOLID REJECT
                    ILLINOIS # 6 RUN 152500 CENTRAX DEASHING, REALISTIC CASE

 YIELD STRUCTURE,              WT% MAF COAL         COMMERCIAL PLANT OUTPUT        COAL TO LIQUEFACTION, T/D AR    33135
                                                     LBS/HR     BBLS/DAY           COAL TO STEAM,        T/D AR     2625
 H2 Consumed, Stage 1              0.0168                                          COAL TO HYDROGEN,     T/D AR     7924
 H2 Consumed, Stage 2              0.0470                                             TOTAL COAL         T/D AR    43685
 Heteroatoms (inc H2O)             0.1498
 C1C3 Gases                        0.0721                                         PLANT CONSTRUCTION COST ($1000)
 C4 - 350F Product                 0.1346                                            LIQUEFACTION                1342755
 350F - 450F Product               0.0716           297767       26150               SOLIDS REMOVAL               287990
 450F - 650F Product               0.2155           158510       12283               HYDROGEN PRODUCTION          658559
 650 - 850F Product                0.2843           476875       34974               BALANCE OF PLANT             468504
    TOTAL C4-850F PRODUCT          0.7061           628992       43000                 TOTAL CONS COST           2757808
 850F+ Product                     0.0000
 Soluble Reject                    0.0568                                         TOTAL CAPITAL REQUIRED         4443490
 Unconverted Coal                  0.0777
 Ash                               0.1300                                         OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                              1562145      116407               COAL COST                    432481
    TOTAL, HYDROTREATED PRODUCT                    1559084      129626               OTHER OPERATING COSTS        374235
                                                                                     BY-PRODUCT CREDITS            97939
 AVERAGE PRODUCT QUALITY                                                               TOTAL,$1000/YR             708777
                         CARBON  HYDROGEN    NITROGEN    SULFUR     OXYGEN    API deg
RAW                       87.57    11.64        0.24      0.10       0.45      22.43    HYDROTREATING COSTS          206619
HYDR                      87.00    13.00                                       40.00      TOTAL INC.HYDROTREAT       915397

 DEASHER COST FACTOR               1.00                                           REQUIRED SELLING PRICE         $/BBL
 COAL CLEANING, %Btu Recovered   100.00 %       0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)            0                                              RAW PRODUCT                   34.30
                                                                                     HYDROTREATED PRODUCT          35.63

                            SG          C            H          N          S          O         API

              NAP         0.7800     85.4100      14.2600     0.0400     0.1800     0.1100     49.91
              MD          0.8840     86.1900      12.4400     0.1300     0.1300     1.1100     28.57
              HHD         0.9340     87.8100      11.5300     0.2500     0.0800     0.3300     20.00
              PHD         1.0020     88.7700      10.2700     0.3600     0.0700     0.5300      9.72
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                  TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250D          28.5300 SOLID REJECT
                           ILLINOIS # 6 RUN TS250D CENTRAX DEASHING, REALISTIC CASE

 YIELD STRUCTURE,           WT% MAF COAL        COMMERCIAL PLANT OUTPUT          COAL TO LIQUEFACTION, T/D AR     33135
                                                 LBS/HR      BBLS/DAY            COAL TO STEAM,        T/D AR      2625
  H2 Consumed, Stage 1         0.0168                                            COAL TO HYDROGEN,     T/D AR      7924
  H2 Consumed, Stage 2         0.0470                                               TOTAL COAL         T/D AR     43685
  Heteroatoms (inc H2O)        0.1498
  C1C3 Gases                   0.0721                                            PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product            0.1346            297767       26150                 LIQUEFACTION                1342755
  350F - 450F Product          0.0716            158510       12283                 SOLIDS REMOVAL               431985
  450F - 650F Product          0.2155            476875       34974                 HYDROGEN PRODUCTION          658559
  650 - 850F Product           0.2843            628992       43000                 BALANCE OF PLANT             468504
     TOTAL C4-850F PRODUCT     0.7061                                                  TOTAL CONS COST          2901804
  850F+ Product                0.0000
  Soluble Reject               0.0568                                            TOTAL CAPITAL REQUIRED         4658358
  Unconverted Coal             0.0777
  Ash                          0.1300                                            OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                           1562145      116407                COAL COST                    432481
     TOTAL, HYDROTREATED PRODUCT                 1559884      129626                OTHER OPERATING COSTS        384135
                                                                                    BY-PRODUCT CREDITS            97939
 AVERAGE PRODUCT QUALITY                                                               TOTAL,$1000/YR            718677
                     CARBON    HYDROGEN    NITROGEN    SULFUR    OXYGEN    API deg
RAW                   87.57     11.64        0.24       0.10      0.45      22.43       HYDROTREATING COSTS          206619
HYDR                  87.00     13.00                                       40.00         TOTAL INC.HYDROTREAT      925296

 DEASHER COST FACTOR            1.50                                             REQUIRED SELLING PRICE           $/BBL
 COAL CLEANING, %Btu Recovered  100.00 %        0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)        0                                                 RAW PRODUCT                   35.32
                                                                                    HYDROTREATED PRODUCT          36.55


                                  SG          C          H          N          S          O         API

                        NAP     0.7800     85.4100    14.2600     0.0400     0.1800     0.1100     49.91
                        MD      0.8840     86.1900    12.4400     0.1300     0.1300     1.1100     28.57
                        MHD     0.9340     87.8100    11.5300     0.2500     0.0800     0.3300     20.00
                        PHD     1.0020     88.7700    10.2700     0.3600     0.0700     0.5300      9.72
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                    TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G          68.0000 SOLID REJECT
                             ILLINOIS # 6 19250G K-M CRITICAL SOLVENT DEASHING

 YIELD STRUCTURE,              WT% MAF COAL        COMMERCIAL PLANT OUTPUT               COAL TO LIQUEFACTION, T/D AR     33557
                                                    LBS/HR     BBLS/DAY                  COAL TO STEAM,        T/D AR      2651
 H2 Consumed, Stage 1            0.0185                                                  COAL TO HYDROGEN,     T/D AR      8032
 H2 Consumed, Stage 2            0.0476                                                     TOTAL COAL         T/D AR     44240
 Heteroatoms (inc H2O)           0.1573
 C1C3 Gases                      0.0780                                                  PLANT CONSTRUCTION COST ($1000)
 C4 - 350F                       0.1534              343700      30736                      LIQUEFACTION                1366197
 350F - 450F Product             0.0775              173642      13609                      SOLIDS REMOVAL               140664
 450F - 650F Product             0.2359              528545      38557                      HYDROGEN PRODUCTION          681705
 650 - 850F Product              0.2162              484402      32951                      BALANCE OF PLANT             464778
    TOTAL C4-850 PRODUCT         0.6830                                                        TOTAL CONS COST          2653344
 850F+ Product                   0.0000
 Soluble Reject                  0.0822                                                  TOTAL CAPITAL REQUIRED         4286873
 Unconverted Coal                0.0655
 Ash                             0.1150                                                  OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                               1530288     115853                     COAL COST                     437980
    TOTAL, HYDROTREATED PRODUCT                     1521682     126452                     OTHER OPERATING COSTS         359917
                                                                                           BY-PRODUCT CREDITS            116421
 AVERAGE PRODUCT QUALITY                                                                     TOTAL,$1000/YR              681476
                    CARBON   HYDROGEN     NITROGEN    SULFUR      OXYGEN      API deg
 I                   87.21     11.60         0.18       0.08        0.93       24.89       HYDROTREATING COSTS           205401
 H                   87.00     13.00                                           40.00          TOTAL INC.HYDROTREAT       886877

 DEASHER COST FACTOR                1.00                                                 REQUIRED SELLING PRICE          $/BBL
 COAL CLEANING, %Btu Recovered    100.00 %       0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)           0                                                     RAW PRODUCT                    33.19
                                                                                           HYDROTREATED PRODUCT           35.33

                      SG          C           H          N          S          O         API

 NAP                0.7660     84.8100     13.8300     0.0500     0.1500     1.1600     53.23
 HD                 0.8740     85.7200     11.9300     0.1400     0.1000     2.1100     30.40
 MID                0.9390     87.4500     11.3200     0.[illegible]900     0.0600     0.9800     19.19
 PHD                1.0070     89.1800     10.1900     0.2800     0.0500     0.3000      9.02
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TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G            90.1400 SOLID REJECT
              ILLINOIS # 6 RUN IS250G CENTRAX DEASHING IDEAL CASE, 1:0.4 FEED:ESTRACTANT

  YIELD STRUCTURE,              #/# MAF COAL         COMMERCIAL PLANT OUTPUT              COAL TO LIQUEFACTION, T/D AR     33557
                                                      LBS/HR      BBLS/DAY                COAL TO STEAM,        T/D AR      2719
  H2 Consumed, Stage 1             0.0186                                                 COAL TO HYDROGEN,     T/D AR     10773
  H2 Consumed, Stage 2             0.0501                                                    TOTAL COAL         T/D AR     47049
  Heteroatoms (inc H2O)            0.1586
  C1C3 Gases                       0.0800                                                 PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product                0.1543                                                    LIQUEFACTION                 1404622
  350F - 450F Product              0.0779             345649       30910                     SOLIDS REMOVAL                172098
  450F - 650F Product              0.2372             174627       13686                     HYDROGEN PRODUCTION           694855
  650 - 850F Product               0.2811             531542       38776                     BALANCE OF PLANT              476908
     TOTAL C4-850F PRODUCT         0.7505             629820       42843                       TOTAL CONS COST            2748482
  850F+ Product                    0.0000
  Soluble Reject                   0.0193                                                 TOTAL CAPITAL REQUIRED          4443420
  Unconverted Coal                 0.0602
  Ash                              0.1158                                                 OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                                                                       COAL COST                     465786
     TOTAL, HYDROTREATED PRODUCT                      1681638      126215                    OTHER OPERATING COSTS         371061
                                                      1675388      139225                    BY-PRODUCT CREDITS            126540
                                                                                               TOTAL,$1000/YR              710307
 AVERAGE PRODUCT QUALITY
                         CARBON  HYDROGEN     NITROGEN     SULFUR      OXYGEN      API deg      HYDROTREATING COSTS        244376
RAW                       87.38    11.48         0.19        0.08        0.88        23.54        TOTAL INC.HYDROTREAT     954682
HYDR                      87.00    13.00                                             40.00
                                                                                          REQUIRED SELLING PRICE           $/BBL
  DEASHER COST FACTOR               1.00
  COAL CLEANING, %Btu Recovered   100.00 %        0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)           0                                                     RAW PRODUCT                    31.67
  No. Centrax Units Reqd           2097                                                     HYDROTREATED PRODUCT           34.03


                                         SG          C           H          N          S          O          API

                     NAP               0.7660     84.8100     13.8300     0.0500     0.1500     1.1600      53.23
                     ND                0.8740     85.7200     11.9300     0.1400     0.1000     2.1100      30.40
                     MID               0.9390     87.4500     11.3200     0.1900     0.0600     0.9800      19.19
                     PHD               1.0070     89.1800     10.1900     0.2800     0.0500     0.3000       9.02
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                 TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G           90.1400 SOLID REJECT
                          ILLINOIS # 6 RUN 15250G CENTRAX DEASHING IDEAL CASE, 1:0.4 FEED:ESTRACTANT

                                                  COMMERCIAL PLANT OUTPUT          COAL TO LIQUEFACTION, T/D AR     33557
 YIELD STRUCTURE,              W/W MAF COAL         LBS/HR      BBLS/DAY           COAL TO STEAM,        T/D AR      2719
                                                                                   COAL TO HYDROGEN,     T/D AR     10773
 H2 Consumed, Stage 1             0.0186                                              TOTAL COAL         T/D AR     47049
 H2 Consumed, Stage 2             0.0501
 Heteroatoms (inc H2O)            0.1586                                           PLANT CONSTRUCTION COST ($1000)
 C1C3 Gases                       0.0800                                              LIQUEFACTION                1404622
 C4 - 350F Product                0.1543            345649       30910                SOLIDS REMOVAL               258147
 350F - 450F Product              0.0779            174627       13686                HYDROGEN PRODUCTION          694855
 450F - 650F Product              0.2372            531542       38776                BALANCE OF PLANT             476908
 650 - 850F Product               0.2811            629820       42843                  TOTAL CONS COST           2834531
    TOTAL C4-850F PRODUCT         0.7505
 850F+ Product                    0.0000                                           TOTAL CAPITAL REQUIRED         4571821
 Soluble Reject                   0.0193
 Unconverted Coal                 0.0602                                           OPERATING COSTS,$1000/YR
 Ash                              0.1158                                              COAL COST                    465786
     TOTAL,RAW PRODUCT                             1681638      126215                OTHER OPERATING COSTS        376977
     TOTAL, HYDROTREATED PRODUCT                   1675388      139225                BY-PRODUCT CREDITS           126540
                                                                                        TOTAL,$1000/YR             716223
   AVERAGE PRODUCT QUALITY
                       CARBON    HYDROGEN     NITROGEN    SULFUR     OXYGEN     API deg    HYDROTREATING COSTS        244376
 RAW                    87.38     11.48         0.19       0.08       0.88      23.54        TOTAL INC.HYDROTREAT     960598
 HYDR                   87.00     13.00                                         40.00
                                                                                   REQUIRED SELLING PRICE           $/BBL
   DEASHER COST FACTOR              1.50
   COAL CLEANING, %Btu Recovered  100.00 %       0 Flag,1 or 0                        RAW PRODUCT                   32.23
   COAL DRYING FLAG (1 or 0)          0                                               HYDROTREATED PRODUCT          34.56
   No. Centrax Units Reqd          2097


                           SG          C            H          N          S          O        API

             NAP         0.7660      84.8100      13.8300    0.0500     0.1500     1.1600     53.23
             HD          0.8740      85.7200      11.9300    0.1400     0.1000     2.1100     30.40
             HHD         0.9390      87.4500      11.3200    0.1900     0.0600     0.9800     19.19
             PHD         1.0070      89.1800      10.1900    0.2800     0.0500     0.3000      9.02
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                    TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G          76.2700 SOLID REJECT
                              ILLINOIS # 6 RUN 15250G CENTRAX DEASHING, OPTIMISTIC CASE, 1:0.4 FEED:EXTRACTANT

                                                 COMMERCIAL PLANT OUTPUT
  YIELD STRUCTURE,             #/# MAF COAL       LBS/HR    BBLS/DAY            COAL TO LIQUEFACTION, T/D AR     33557
                                                                                COAL TO STEAM,        T/D AR      2719
  H2 Consumed, Stage 1            0.0186                                        COAL TO HYDROGEN,     T/D AR      9367
  H2 Consumed, Stage 2            0.0488                                           TOTAL COAL         T/D AR     45643
  Heteroatoms (inc H2O)           0.1584
  C1C3 Gases                      0.0791                                        PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product               0.1543          345649     30910                 LIQUEFACTION                1383288
  350F - 450F Product             0.0779          174627     13686                 SOLIDS REMOVAL               185137
  450F - 650F Product             0.2372          531542     38776                 HYDROGEN PRODUCTION          688202
  650 - 850F Product              0.2452          549493     37379                 BALANCE OF PLANT             472214
     TOTAL C4-850F PRODUCT        0.7147                                              TOTAL CONS COST          2728841
  850F+ Product                   0.0000
  Soluble Reject                  0.0548                                        TOTAL CAPITAL REQUIRED         4407907
  Unconverted Coal                0.0602
  Ash                             0.1158                                        OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                            1601311    120751                COAL COST                    451848
     TOTAL, HYDROTREATED PRODUCT                  1593707    132437                OTHER OPERATING COSTS        369838
                                                                                   BY-PRODUCT CREDITS           121796
 AVERAGE PRODUCT QUALITY                                                              TOTAL,$1000/YR            699909
                     CARBON   HYDROGEN    NITROGEN    SULFUR     OXYGEN    API deg
RAW                   87.29     11.54        0.19       0.08       0.91      24.27        HYDROTREATING COSTS     223088
HYDR                  87.00     13.00                                        40.00           TOTAL INC.HYDROTREAT 922997

  DEASHER COST FACTOR             1.00                                          REQUIRED SELLING PRICE          $/BBL
  COAL CLEANING, %Btu Recovered 100.00 %        0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)          0                                             RAW PRODUCT                   32.72
  No. Centrax Units Reqd          2312                                             HYDROTREATED PRODUCT          34.94


                                   SG         C          H         N         S         O        API

                      NAP        0.7660    84.8100    13.8300    0.0500    0.1500    1.1600    53.23
                      HD         0.8740    85.7200    11.9300    0.1400    0.1000    2.1100    30.40
                      MHD        0.9390    87.4500    11.3200    0.1900    0.0600    0.9800    19.19
                      PHD        1.0070    89.1800    10.1900    0.2800    0.0500    0.3000     9.02
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TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G          76.2700 SOLID REJECT
            ILLINOIS # 6 RUN 19250G CENTRAX DEASHING, OPTIMISTIC CASE, 1:0.4 FEED:EXTRACTANT

 YIELD STRUCTURE,              #/# MAF COAL          COMMERCIAL PLANT OUTPUT            COAL TO LIQUEFACTION, T/D AR     33557
                                                       LBS/HR     BBLS/DAY              COAL TO STEAM,        T/D AR      2719
  H2 Consumed, Stage 1            0.0186                                                COAL TO HYDROGEN,     T/D AR      9367
  H2 Consumed, Stage 2            0.0488                                                   TOTAL COAL         T/D AR     45643
  Heteroatoms (inc H2O)           0.1584
  C1C3 Gases                      0.0791                                                PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product               0.1543                                                   LIQUEFACTION                1383288
  350F - 450F Product             0.0779           345649       30910                      SOLIDS REMOVAL               277706
  450F - 650F Product             0.2372           174627       13686                      HYDROGEN PRODUCTION          688202
  650 - 850F Product              0.2452           531542       38776                      BALANCE OF PLANT             472214
     TOTAL C4-850F PRODUCT        0.7147           549493       37379                         TOTAL CONS COST           2821410
  850F+ Product                   0.0000
  Soluble Reject                  0.0548                                                TOTAL CAPITAL REQUIRED          4546037
  Unconverted Coal                0.0602
  Ash                             0.1150                                                OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                                                                     COAL COST                    451868
     TOTAL, HYDROTREATED PRODUCT                   1601311      120751                     OTHER OPERATING COSTS        376202
                                                   1593707      132437                     BY-PRODUCT CREDITS           121796
 AVERAGE PRODUCT QUALITY                                                                      TOTAL,$1000/YR            706274
                    CARBON     HYDROGEN      NITROGEN     SULFUR      OXYGEN      API deg
RAW                  87.29       11.54         0.19        0.00        0.91       24.27     HYDROTREATING COSTS         223088
HYDR                 87.00       13.00                                            40.00       TOTAL INC.HYDROTREAT      929362

 DEASHER COST FACTOR              1.50                                                  REQUIRED SELLING PRICE           $/BBL
 COAL CLEANING, %Btu Recovered  100.00 %         0 Flag,1 or 0
 CUAL DRYING FLAG (1 or 0)          0                                                     RAW PRODUCT                    33.35
 No. Centrax Units Reqd          2312                                                     HYDROTREATED PRODUCT           35.52


                        SG          C           H          N          S          O         API

          NAP         0.7660     84.8100     13.8300     0.0500     0.1500     1.1600     53.23
          MD          0.8740     85.7200     11.9300     0.1400     0.1000     2.1100     30.40
          HVD         0.9390     87.4500     11.3200     0.1900     0.0600     0.9800     19.13
          PHD         1.0070     89.1800     10.1900     0.2800     0.0500     0.3000      9.02
          TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G          66.4000 SOLID REJECT
                     ILLINOIS # 6 RUN 15250G CENTRAX DEASHING, REALISTIC CASE. 1:0.4 FEED:EXTRACTANT RATIO

 YIELD STRUCTURE,            #/# MAF COAL          COMMERCIAL PLANT OUTPUT             COAL TO LIQUEFACTION, T/D AR    33557
                                                    LBS/HR      BBLS/DAY               COAL TO STEAM,        T/D AR     2719
 H2 Consumed, Stage 1          0.0186                                                  COAL TO HYDROGEN,     T/D AR     8009
 H2 Consumed, Stage 2          0.0476                                                     TOTAL COAL         T/D AR    44285
 Heteroatoms (inc H2O)         0.1581
 C1C3 Gases                    0.0783                                                  PLANT CONSTRUCTION COST ($1000)
 C4 - 350F Product             0.1543                                                     LIQUEFACTION                1362518
 350F - 450F Product           0.0779               345649       30910                    SOLIDS REMOVAL               185137
 450F - 650F Product           0.2372               174627       13686                    HYDROGEN PRODUCTION          681755
 650 - 850F Product            0.2106               531542       38776                    BALANCE OF PLANT             467626
    TOTAL C4-850F PRODUCT      0.6801               471895       37109                      TOTAL CONS COST           2697036
 850F+ Product                 0.0000
 Soluble Reject                0.0891                                                  TOTAL CAPITAL REQUIRED         4354272
 Unconverted Coal              0.0602
 Ash                           0.1158                                                  OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                              1523713      115472                    COAL COST                    438423
    TOTAL, HYDROTREATED PRODUCT                    1514801      125880                    OTHER OPERATING COSTS        367375
                                                                                          BY-PRODUCT CREDITS           117214
 AVERAGE PRODUCT QUALITY                                                                    TOTAL,$1000/YR             688584
                   CARBON   HYDROGEN     NITROGEN     SULFUR      OXYGEN      API deg
 RAW                87.19    11.61         0.18         0.06        0.94        25.05       HYDROTREATING COSTS          202524
 HYDR               87.00    13.00                                              40.00         TOTAL INC.HYDROTREAT       891107

 DEASHER COST FACTOR            1.00                                                    REQUIRED SELLING PRICE         $/BBL
 COAL CLEANING, %Btu Recovered  100.00 %       0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)         0                                                       RAW PRODUCT                  33.72
 No. Centrax Units Reqd         2312                                                       HYDROTREATED PRODUCT         35.81


                            SG          C            H          N          S          O         API

              NAP         0.7660      84.8100      13.8300    0.0500     0.1500     1.1600      53.23
              MD          0.8740      85.7200      11.9300    0.1400     0.1000     2.1100      30.40
              HHD         0.9390      87.4500      11.3200    0.1900     0.0600     0.9800      19.19
              PHD         1.0070      89.1800      10.1900    0.2800     0.0500     0.3000       9.02
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                TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G          46.4000 SOLID REJECT
                          ILLINOIS # 6 RUN 15250G CENTRAX DEASHING, REALISTIC CASE. 1:0.4 FEED:EXTRACTANT RATIO

 YIELD STRUCTURE,              B/B MAF COAL         COMMERCIAL PLANT OUTPUT         COAL TO LIQUEFACTION, T/D AR     33557
                                                      LBS/HR     BBLS/DAY           COAL TO STEAM,        T/D AR      2719
 H2 Consumed, Stage 1            0.0186                                             COAL TO HYDROGEN,     T/D AR      8009
 H2 Consumed, Stage 2            0.0476                                                 TOTAL COAL        T/D AR     44285
 Heteroatoms (inc H2O)           0.1501
 C1C3 Gases                      0.0783                                             PLANT CONSTRUCTION COST ($1000)
 C4 - 350F Product               0.1543             345649      30910                   LIQUEFACTION                1362518
 350F - 450F Product             0.0779             174627      13686                   SOLIDS REMOVAL               277706
 450F - 650F Product             0.2372             531542      38776                   HYDROGEN PRODUCTION          681755
 650 - 850F Product              0.2106             471095      32100                   BALANCE OF PLANT             467626
    TOTAL C4-850F PRODUCT        0.6801                                                   TOTAL CONS COST           2789604
 850F+ Product                   0.0000
 Soluble Reject                  0.0891                                             TOTAL CAPITAL REQUIRED          4492402
 Unconverted Coal                0.0602
 Ash                             0.1158                                             OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                                                                 COAL COST                     438423
     TOTAL, HYDROTREATED PRODUCT                   1523713     115472                  OTHER OPERATING COSTS         373740
                                                   1514801     125880                  BY-PRODUCT CREDITS            117214
 AVERAGE PRODUCT QUALITY                                                                  TOTAL,$1000/YR             694948
                         CARBON  HYDROGEN   NITROGEN   SULFUR    OXYGEN   API deg
RAW                       87.19    11.61       0.10      0.00      0.94     25.05      HYDROTREATING COSTS           202524
HYDR                      87.00    13.00                                    40.00        TOTAL INC.HYDROTREAT        897471

 DEASHER COST FACTOR               1.50                                             REQUIRED SELLING PRICE          $/BBL
 COAL CLEANING, %Btu Recovered   100.00 %       0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)            0                                                RAW PRODUCT                   34.39
 No. Centrax Units Reqd            2312                                                HYDROTREATED PRODUCT          36.42


                                       SG          C          H          N        S         O        API

                        NAP          0.7660     84.0100    13.8300    0.0500    0.1500    1.1600    53.23
                        MD           0.8740     85.7200    11.9300    0.1400    0.1000    2.1100    30.40
                        HVD          0.9390     87.4500    11.3200    0.1900    0.0600    0.9800    19.19
                        PHD          1.0070     89.1000    10.1900    0.2000    0.0500    0.3000     9.02
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TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-   250G           96.2600 SOLID REJECT
         ILLINOIS # 6 RUN IS250G CENTRAX DEASHING IDEAL CASE, 1:0.7 FEED:EXTRACTANT

YIELD STRUCTURE,              #/# MAF COAL          COMMERCIAL PLANT OUTPUT        COAL TO LIQUEFACTION, T/D AR     33557
                                                     LBS/HR     BBLS/DAY           COAL TO STEAM,        T/D AR      2730
 H2 Consumed, Stage 1             0.0186                                           COAL TO HYDROGEN,     T/D AR     11264
 H2 Consumed, Stage 2             0.0505                                              TOTAL COAL         T/D AR     47551
 Heteroatoms (inc H20)            0.1587
 C1C3 Gases                       0.0803                                           PLANT CONSTRUCTION COST ($1000)
 C4 - 350F Product                0.1543                                              LIQUEFACTION                1412044
 350F - 450F Product              0.0779          345649      30910                   SOLIDS REMOVAL               205754
 450F - 650F Product              0.2372          174627      13686                   HYDROGEN PRODUCTION          696997
 650 - 850F Product               0.2936          531542      38776                   BALANCE OF PLANT             479059
    TOTAL C4-850F PRODUCT         0.7631          657903      44753                     TOTAL CONS COST           2793854
 850F+ Product                    0.0000
 Soluble Reject                   0.0068                                           TOTAL CAPITAL REQUIRED         4513891
 Unconverted Coal                 0.0602
 Ash                              0.1158                                           OPERATING COSTS,$1000/YR
    TOTAL,RAW PRODUCT                            1709721     128125                   COAL COST                    470759
    TOTAL, HYDROTREATED PRODUCT                  1703944     141598                   OTHER OPERATING COSTS        375357
                                                                                      BY-PRODUCT CREDITS           128208
 AVERAGE PRODUCT QUALITY                                                                TOTAL,$1000/YR             717908
                      CARBON   HYDROGEN   NITROGEN   SULFUR    OXYGEN    API deg
RAW                    87.41     11.45       0.19      0.08      0.87      23.30      HYDROTREATING COSTS          251818
HYDR                   87.00     13.00                                     40.00        TOTAL INC.HYDROTREAT       969726

 DEASHER COST FACTOR              1.00                                             REQUIRED SELLING PRICE          $/BBL
 COAL CLEANING, %Btu Recovered  100.00 %      0 Flag,1 or 0
 COAL DRYING FLAG (1 or 0)           0                                                RAW PRODUCT                   31.61
 No. Centrax Units Reqd           2656                                                HYDROTREATED PRODUCT          33.99


                                 SG          C           H          N          S          O         API

                 NAP           0.7660     84.8100     13.8300     0.0500     0.1500     1.1600      53.23
                 MD            0.8740     85.7200     11.9300     0.1400     0.1000     2.1100      30.40
                 HHD           0.9390     87.4500     11.3200     0.1900     0.0600     0.9800      19.19
                 PHD           1.0070     89.1000     10.1900     0.2800     0.0500     0.3000       9.02
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                       TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    2506          96.2600 SOLID REJECT
                                 ILLINOIS # 6 RUN 152506 CENTRAX DEASHING IDEAL CASE, 1:0.7 FEED:EXTRACTANT

                                                    COMMERCIAL PLANT OUTPUT
   YIELD STRUCTURE,               #/# MAF COAL       LBS/HR     BBLS/DAY                  COAL TO LIQUEFACTION, T/D AR     33557
                                                                                          COAL TO STEAM,        T/D AR      2730
   H2 Consumed, Stage 1             0.0186                                                COAL TO HYDROGEN,     T/D AR     11264
   H2 Consumed, Stage 2             0.0505                                                   TOTAL COAL         T/D AR     47551
   Heteroatoms (inc H20)            0.1587
   C1C3 Gases                       0.0803                                                PLANT CONSTRUCTION COST ($1000)
   C4 - 350F Product                0.1543        345649       30910                         LIQUEFACTION               1412044
   350F - 450F Product              0.0779        174627       13686                         SOLIDS REMOVAL              308631
   450F - 650F Product              0.2372        531542       38776                         HYDROGEN PRODUCTION         696997
   650 - 850F Product               0.2956        657903       44753                         BALANCE OF PLANT            479059
      TOTAL C4-850F PRODUCT         0.7631                                                     TOTAL CONS COST          2896731
   850F+ Product                    0.0000
   Soluble Reject                   0.0068                                                TOTAL CAPITAL REQUIRED        4667403
   Unconverted Coal                 0.0602
   Ash                              0.1158                                                OPERATING COSTS,$1000/YR
      TOTAL,RAW PRODUCT                          1709721      128125                         COAL COST                   470759
      TOTAL, HYDROTREATED PRODUCT                1703944      141598                         OTHER OPERATING COSTS       382430
                                                                                             BY-PRODUCT CREDITS          128208
  AVERAGE PRODUCT QUALITY                                                                      TOTAL,$1000/YR            724981
                        CARBON   HYDROGEN      NITROGEN     SULFUR       OXYGEN     API deg
RAW                      87.41     11.45          0.19        0.08         0.87       23.30       HYDROTREATING COSTS         251818
HYDR                     87.00     13.00                                              40.00         TOTAL INC.HYDROTREAT      976799

   DEASHER COST FACTOR               1.50                                                 REQUIRED SELLING PRICE        $/BBL
   COAL CLEANING, %Btu Recovered   100.00 %          0 Flag,1 or 0
   COAL DRYING FLAG (1 or 0)           0                                                     RAW PRODUCT                  32.27
   No. Centrax Units Reqd           2656                                                     HYDROTREATED PRODUCT         36.59


                                   SG           C             H          N         S          O         API

                 NAP             0.7660      84.8100       13.8300    0.0500    0.1500     1.1600     53.23
                 HD              0.8740      85.7200       11.9300    0.1400    0.1000     2.1100     30.40
                 MID             0.9390      87.4500       11.3200    0.1900    0.0600     0.9800     19.19
                 PHD             1.0070      89.1800       10.1900    0.2000    0.0500     0.3000      9.02
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TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G           89.4900 SOLID REJECT
             ILLINOIS # 6 RUN 15250G CENTRAX DEASHING, OPTIMISTIC CASE, 1:0.7 FEED:EXTRACTANT

                                                    COMMERCIAL PLANT OUTPUT
  YIELD STRUCTURE,                  #/# MAF COAL                                               COAL TO LIQUEFACTION, T/D AR    33557
                                                     LBS/HR     BBLS/DAY                       COAL TO STEAM,        T/D AR     2730
  H2 Consumed, Stage 1                0.0186                                                   COAL TO HYDROGEN,     T/D AR    10716
  H2 Consumed, Stage 2                0.0500                                                      TOTAL COAL         T/D AR    47004
  Heteroatoms (inc H2O)               0.1586
  C1C3 Gases                          0.0799                                                   PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product                   0.1543                                                      LIQUEFACTION                1403773
  350F - 450F Product                 0.0779            345649      30910                         SOLIDS REMOVAL               218561
  450F - 650F Product                 0.2372            174627      15686                         HYDROGEN PRODUCTION          694412
  650 - 850F Product                  0.2797            531542      38776                         BALANCE OF PLANT             477244
     TOTAL C4-850F PRODUCT            0.7491            626612      42625                           TOTAL CONS COST           2793989
  850F+ Product                       0.0000
  Soluble Reject                      0.0207                                                   TOTAL CAPITAL REQUIRED         4511790
  Unconverted Coal                    0.0402
  Ash                                 0.1158                                                   OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                                 1678430     125997                         COAL COST                    465337
     TOTAL, HYDROTREATED PRODUCT                       1672125     138954                         OTHER OPERATING COSTS        375670
                                                                                                  BY-PRODUCT CREDITS           126360
  AVERAGE PRODUCT QUALITY                                                                           TOTAL,$1000/YR             714647
                           CARBON   HYDROGEN   NITROGEN    SULFUR     OXYGEN     API deg
RAW                         87.37     11.40       0.19      0.08       0.88       23.57            HYDROTREATING COSTS          243525
HYDR                        87.00     13.00                                       40.00              TOTAL INC.HYDROTREAT       958173

  DEASHER COST FACTOR                  1.00                                                    REQUIRED SELLING PRICE          $/BBL
  COAL CLEANING, %Btu Recovered      100.00 %      0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)              0                                                        RAW PRODUCT                  32.05
  No. Centrax Units Reqd               2872                                                       HYDROTREATED PRODUCT         34.38


                                SG          C          H          N          S          O         API

               NAP            0.7660     84.8100    13.8300     0.0500     0.1500     1.1600     53.23
               ND             0.8740     85.7200    11.9300     0.1400     0.1000     2.1100     30.40
               MID            0.9390     87.4500    11.3200     0.1900     0.0600     0.9800     19.19
               PHD            1.0070     89.1800    10.1900     0.2800     0.0500     0.3000      9.02
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                TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G          89.4900 SOLID REJECT
                          ILLINOIS # 6 RUN TS250G CENTRAX DEASHING, OPTIMISTIC CASE, 1:0.7 FEED:EXTRACTANT

                                            COMMERCIAL PLANT OUTPUT
  YIELD STRUCTURE,          #/# MAF COAL       LBS/HR      BBLS/DAY                COAL TO LIQUEFACTION, T/D AR    33557
                                                                                   COAL TO STEAM,        T/D AR     2730
  H2 Consumed, Stage 1        0.0186                                               COAL TO HYDROGEN,     T/D AR    10716
  H2 Consumed, Stage 2        0.0500                                                  TOTAL COAL         T/D AR    47004
  Heteroatoms (inc H2O)       0.1586
  C1C3 Gases                  0.0799                                               PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product           0.1543                                                  LIQUEFACTION                1403773
  350F - 450F Product         0.0779            345649        30910                   SOLIDS REMOVAL               327841
  450F - 650F Product         0.2372            174627        13686                   HYDROGEN PRODUCTION          694412
  650 - 850F Product          0.2797            531542        38776                   BALANCE OF PLANT             477244
     TOTAL C4-850F PRODUCT    0.7491            626612        42625                      TOTAL CONS COST          2903270
  850F+ Product               0.0000
  Soluble Reject              0.0207                                               TOTAL CAPITAL REQUIRED          4674857
  Unconverted Coal            0.0602
  Ash                         0.1158                                               OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                                                                COAL COST                    465337
     TOTAL, HYDROTREATED PRODUCT                1678430       125997                  OTHER OPERATING COSTS        383183
                                                1672125       138954                  BY-PRODUCT CREDITS           126360
  AVERAGE PRODUCT QUALITY                                                                TOTAL,$1000/YR            722160
                       CARBON   HYDROGEN    NITROGEN    SULFUR     OXYGEN     API deg
RAW                     87.37    11.48        0.19       0.08       0.88       23.57        HYDROTREATING COSTS          243525
HYDR                    87.00    13.00                                         40.00          TOTAL INC.HYDROTREAT       965686

  DEASHER COST FACTOR           1.50                                               REQUIRED SELLING PRICE          $/BBL
  COAL CLEANING, %Btu Recovered 100.00 %     0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)        0                                                  RAW PRODUCT                  32.77
  No. Centrax Units Reqd        2872                                                  HYDROTREATED PRODUCT         35.03


                              SG          C          H          N          S          O          API

            NAP             0.7660     84.8100    13.8300    0.0500     0.1500     1.1600      53.23
            MD              0.8740     85.7200    11.9300    0.1400     0.1000     2.1100      30.40
            HVD             0.9390     87.4500    11.3200    0.1900     0.0600     0.9000      19.19
            PHD             1.0070     89.1800    10.1900    0.2800     0.0500     0.3000       9.02
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TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G           81.4900 SOLID REJECT
        ILLINOIS # 6 RUN IS250G CENTRAX DEASHING, REALISTIC CASE. 1:0.7 FEED:EXTRACTANT RATIO

                                            COMMERCIAL PLANT OUTPUT            COAL TO LIQUEFACTION, T/D AR    33557
YIELD STRUCTURE,          #/# MAF COAL        LBS/HR     BBLS/DAY              COAL TO STEAM,        T/D AR     2730
                                                                               COAL TO HYDROGEN,     T/D AR     9952
H2 Consumed, Stage 1        0.0186                                                TOTAL COAL         T/D AR    46239
H2 Consumed, Stage 2        0.0493
Heteroatoms (inc H2O)       0.1585                                             PLANT CONSTRUCTION COST ($1000)
C1C3 Gases                  0.0795                                                LIQUEFACTION                1392188
C4 - 350F Product           0.1543            345649      30910                   SOLIDS REMOVAL               218561
350F - 450F Product         0.0779            174627      13686                   HYDROGEN PRODUCTION          690798
450F - 650F Product         0.2372            531542      38776                   BALANCE OF PLANT             474697
650 - 850F Product          0.2602            582933      39653                     TOTAL CONS COST           2776244
   TOTAL C4-850F PRODUCT    0.7296
850F+ Product               0.0000                                             TOTAL CAPITAL REQUIRED         4481835
Soluble Reject              0.0400
Unconverted Coal            0.0602                                             OPERATING COSTS,$1000/YR
Ash                         0.1158                                                COAL COST                    457769
    TOTAL,RAW PRODUCT                        1634751     123026                   OTHER OPERATING COSTS        374295
    TOTAL, HYDROTREATED PRODUCT              1627711     135263                   BY-PRODUCT CREDITS           123781
                                                                                     TOTAL,$1000/YR            708283
  AVERAGE PRODUCT QUALITY
                        CARBON   HYDROGEN    NITROGEN    SULFUR     OXYGEN    API deg    HYDROTREATING COSTS          231950
RAW                      87.32     11.51       0.19       0.08       0.90      23.96       TOTAL INC.HYDROTREAT       940233
HYDR                     87.00     13.00                                       40.00
                                                                               REQUIRED SELLING PRICE          $/BBL
   DEASHER COST FACTOR             1.00
   COAL CLEANING, %Btu Recovered 100.00 %      0 Flag,1 or 0                      RAW PRODUCT                   32.57
   COAL DRYING FLAG (1 or 0)          0                                           HYDROTREATED PRODUCT          34.82
   No. Centrax Units Reqd          2872


                           SG         C          H          N        S         O        API

              NAP        0.7660    84.8100    13.8300    0.0500   0.1500    1.1600    53.23
              HD         0.8740    85.7200    11.9300    0.1400   0.1000    2.1100    30.40
              MID        0.9390    87.4500    11.3200    0.1900   0.0600    0.9800    19.19
              PHD        1.0070    89.1800    10.1900    0.2800   0.0500    0.3000     9.02
```

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               TWO STAGE LIQUEFACTION TECHNO/ECONOMIC MODEL-RUN-    250G            81.4900 SOLID REJECT
                        ILLINOIS # 6 RUN 15250G CENTRAX DEASHING, REALISTIC CASE. 1:0.7 FEED:EXTRACTANT RATIO

  YIELD STRUCTURE,              #/# MAF COAL          COMMERCIAL PLANT OUTPUT          COAL TO LIQUEFACTION, T/D AR     33557
                                                      LBS/HR       BBLS/DAY             COAL TO STEAM,        T/D AR      2730
  H2 Consumed, Stage 1            0.0186                                               COAL TO HYDROGEN,     T/D AR      9952
  H2 Consumed, Stage 2            0.0493                                                  TOTAL COAL         T/D AR     46239
  Heteroatoms (inc H2O)           0.1585
  C1C3 Gases                      0.0795                                               PLANT CONSTRUCTION COST ($1000)
  C4 - 350F Product               0.1543                                                  LIQUEFACTION                 1392188
  350F - 450F Product             0.0779              345649        30910                 SOLIDS REMOVAL                327841
  450F - 650F Product             0.2372              174627        13686                 HYDROGEN PRODUCTION           690798
  650 - 850F Product              0.2602              531542        38776                 BALANCE OF PLANT              474697
     TOTAL C4-850F PRODUCT        0.7296              582933        39653                    TOTAL CONS COST           2885524
  850F+ Product                   0.0000
  Soluble Reject                  0.0400                                               TOTAL CAPITAL REQUIRED          4644902
  Unconverted Coal                0.0602
  Ash                             0.1158                                               OPERATING COSTS,$1000/YR
     TOTAL,RAW PRODUCT                                                                    COAL COST                     457769
     TOTAL, HYDROTREATED PRODUCT                      1634751       123026                OTHER OPERATING COSTS         381808
                                                      1627711       135263                BY-PRODUCT CREDITS            123781
  AVERAGE PRODUCT QUALITY                                                                    TOTAL,$1000/YR             715796
                      CARBON    HYDROGEN     NITROGEN    SULFUR     OXYGEN     API deg
RAW                    87.32     11.51          0.19       0.08       0.90       23.96     HYDROTREATING COSTS           231950
HYDR                   87.00     13.00                                           40.00        TOTAL INC.HYDROTREAT        947746

  DEASHER COST FACTOR             1.50                                                 REQUIRED SELLING PRICE           $/BBL
  COAL CLEANING, %Btu Recovered  100.00 %        0 Flag,1 or 0
  COAL DRYING FLAG (1 or 0)          0                                                    RAW PRODUCT                     33.31
  No. Centrax Units Reqd          2072                                                    HYDROTREATED PRODUCT            35.49


                             SG          C            H          N         S          O        API

                NAP        0.7660     84.8100     13.8300    0.0500    0.1500    1.1600      53.23
                KD         0.8740     85.7200     11.9300    0.1400    0.1000    2.1100      30.40
                MID        0.9390     87.4500     11.3200    0.1900    0.0600    0.9800      19.19
                PHD        1.0070     89.1800     10.1900    0.2800    0.0500    0.3000       9.02
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APPENDIX B

APPENDIX B

MITRE

July 16, 1985
W53-253

Dr. James F. Kelly
Head, Coal Liquefaction Section
Synthetic Fuels
Energy Research Laboratories
Canada Center for Mineral and Energy Technology
555 Booth Street
Ottawa, Ontario K1A0G1

Dear Dr. Kelly:

As an interim report to our CENTRAX Process Evaluation study, we are pleased to attach a review paper on the "State-of-the-Art of Coal Extract Deashing Techniques." This paper will eventually be included as a part of our final report on the CENTRAX process. It provides a general background against which the CENTRAX process will be evaluated. Further effort on the CENTRAX process Evaluation project is still pending the receipt of the necessary data requested in our letter dated April 4, 1985, to Sandwell Beak Research Group.

Hoping that you will find our review paper interesting, we remain

Yours sincerely,

Martin B Neuworth

Martin B. Neuworth
Chief Engineer
Energy, Resource and
Environmental Systems Division

MBN:AES:dn

Attachment

bcc: N. Coates
M. Neuworth

The MITRE Corporation
Metrek Division
1820 Dolley Madison Boulevard, McLean, Virginia 22102-3481
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STATE-OF-THE-ART OF COAL LIQUID DEASHING TECHNOLOGY

Solid-liquid separation is an essential unit operation in direct coal liquefaction processes. In this unit operation, the solid unconverted coal and ash are separated from the coal extract. The extent to which these fine solids (usually less than 10 microns) must be removed is dictated by the desired level of coal liquid recovery, the required liquid product quality, or specific end-user need, the economics of downstream refining of the recovered coal liquid products, and the internal requirements of the process for a recycle fluid stream.

To date, a number of solid-liquid separation techniques have been applied for deashing crude coal liquefaction products. These include the commercially established vacuum distillation technique, some conventional mechanical separation techniques such as filtration, hydrocloning, and centrifugation, as well as some newly introduced techniques like critical solvent deashing and anti-solvent deashing which are still under development.

Distillation

Distillation is the process of evaporating part of the liquid in a slurry to give at least one overhead, solid-free liquid stream and a bottom stream containing some liquid and the undissolved solids. Vacuum distillation permits the operating temperature to be reduced substantially, but the column diameter and energy inputs are increased. A substantial amount of residual oil must be left in the vacuum still bottoms to maintain its fluidity and make it pumpable for further processing. The exact amount of residual oil left depends upon the liquid viscosity. Processes which produce distillable low-viscosity liquids are, therefore, more amenable to distillation.

Vacuum distillation is thus limited in applicability to the deashing of syncrudes which are the main raw liquid products from the single-stage direct coal liquefaction processes like SRC-II, Exxon Donor Solvent (EDS), and H-Coal. In these processes, vacuum distillation has been successfully used to recover the distillable oils of the syncrude, leaving a pumpable vacuum-still bottoms residue consisting of as much as 50-percent unconverted coal and ash admixed with an equal percentage of undistilled material.

Filtration

Filtration is a process of physically screening solid particles from a slurry. The particles are trapped over the filter medium while the liquid is allowed to pass through and is collected as a filtrate. Filtration has been applied with little success for the deashing of nondistillable residual liquids, produced as the major, intermediate, or final conversion product of such coal liquefaction processes as SRC-I, Integrated Two-Stage Liquefaction, the British Liquid Solvent Extraction, and Supercritical Solvent Coal Extraction.

In the past few years, filtration of SRC-I was evaluated in the United States at both Fort Lewis and Wilsonville, at 50 tons/day and 6 tons/day scale, respectively.(1) The experience using this technique confirmed the problems encountered earlier by the Germans during World War II in the Pott-Broche process, where candle-type pressure ceramic filters were used.(2) These problems include:

- Low filtration rate per unit cross-sectional area
- long cycle times of more than three hours
- Limited scale-up capability
- High sensitivity of filtration rate to changes in the suspended solid particle size distribution of the filter feed
- High liquid holdup, of as much as 50 percent of the coal liquid product, may be retained in the residual filter cake
- Redundancy in pumps, valves, and other operating units is necessary for smooth operation, resulting in high capital and maintenance costs.

Table 2-1 contains a summary of the major filtration experiences, other than those of Fort Lewis and Wilsonville, in coal liquefaction processes.(3) Also listed, where available, is the estimated filtrate flow rate required for filtration to be economically feasible, and a summary of the solids removal capability of precoat filtration. It is evident that filtration gives a very sharp separation with a high degree of solids removal from the liquid phase. Since filtration generally works well with high solids concentration, it would be preferred to vacuum distillation for coal extracts with high mineral matter concentration. However, the filtrate flow rates are typically low for filtration to be considered an economical coal extract deashing technique. The addition of filter aids to increase the filtration

TABLE 2-1

SUMMARY OF MAJOR FILTRATION EXPERIENCE IN COAL LIQUEFACTION PROCESSES

Process	Feed Descrip.	Temp. (°F)	Pressure (psig)	Filtrate Flowrate (lb/hr/ft²)	Pressure Drop (psi)	Solids in Feed (wt%)	Solids in Filtrate (wt%)	Equivalent Solids in Product (wt%)	Equipment Description	Precoat Removal Rate (in/hr)	Filtrate Flowrate Required to be Economical (lb/hr/ft x ft)
SRC(17) (PAMCO)	Dissolver Product	456		22	22.5	6.0	0.031-0.040	0.12-	Goslin-Birmingham, 12 in. wide x 23 in. diameter, rotary precoat filter. Shell heated. 20 x 250 Dutch weave stainless steel filter cloth. Supercel precoat. 7.5% Fiber Flo in base.	0.030	
H-Coal	Reactor Effluent (Distillate Crude or Fuel Oil) with a light oil added. Viscosity = 1-2 cp.	600	30	1607	15	30.	<0.10		Goslin-Birmingham, 12 in. wide x 18 in. diameter, rotary precoat filter. Precoat thickness = 3 in.	0.240-0.360	
CSF(10)	Extract made with start-up solvent or natural solvent @ 710-750°F, residence time = 20 min.-1 hr.	600	~150	716 (based on submerged area)	7.4	8.3	0.08	0.33 (ash)	Dorrco, rotary precoat filter. 30 x 150 mesh plain twill, stainless steel cloth. Precoat-Celite 503.	0.061	~300 (submerged area)
COED (37)	Ill. #6 coal-derived oil.	345	10.0	63	10.0	7.5	<0.1	<0.1	Goslin-Birmingham, 13 inc. wide x 3 ft. diameter, rotary precoat filter. 24 x 250 Dutch weave. Precoat-Celite 503.	0.060	~70

Source: Reference (3)

rate does not appear to be useful because of the amount of filter aid required, combined with further reduction in liquid recovery due to the additional liquid retained by the expended cake.

2.3 Hydroclones

Hydroclones, or liquid cyclones, employ centrifugal force to separate solid particles from a flowing liquid (slurry) stream. Hydroclones have no moving parts and are essentially trouble-free in operation, even at high throughput. As such, hydroclones represent the least expensive deashing means for coal liquid extracts. However, hydroclones proved ineffective for removing fine solid particles with diameters of less than 7 microns. Hydroclones have, therefore, proved ineffective in separation of solids in the two- to five-micron range, which is the typical size range of fine suspended solids in coal liquid extracts. Hydroclones also become ineffective at viscosities greater than about 30 centipoises, which is not unusual for coal liquid extracts.

The major studies conducted to date on the evaluation of hydroclones as a means of separating insoluble materials (solids) from coal-derived extracts have been carried out by Hydrocarbon Research, Inc. (HRI)(3), and by Consolidation Coal Co. (CONSOL).(3) Hydroclones are an essential unit operation in the syncrude mode of the H-Coal process. The small-scale hydroclone separation tests were made with one unit of a Dorr-Oliver six-cyclone, "DOXIE."(3) The results indicated that the clean liquid will have about one-third of the solids content of the underflow, with a coal liquid extract feed specific gravity of 0.9 and viscosity of 3.5 centipoises.

CONSOL examined a coal liquid extract deashing system consisting of two 3-inch primary hydroclones in parallel, followed by a single 3-inch wash hydroclone. This system resulted in solids removal of 93.9 to 95.8 weight percent associated with an extract recovery of 90.1 to 96.7 weight percent.(3) However, the liquid overflow from this system had a total solids content of 2.03 to 3.34 weight percent, compared to a value of 0.8 weight percent obtained by filtration, on the same feed.(3)

The main disadvantages of the hydroclone system when compared to filtration as a coal liquid deashing technique are the lower solids removal efficiency and the higher residual liquid content in the rejected solids. In their most recent evaluation,(3) Foster-Wheeler concluded that hycroclones are not the most efficient solids removal method, even when two stages are used on the coal extract. It appears that very fine particles (two to five microns) in the coal extract, restrict product specification for the clean liquid product, and cost considerations will require the use of alternate solids separation means, either in place of, or in conjunction with hydroclones, in most coal liquefaction processes.

Centrifuges

Centrifuging is the process of subjecting the solids-containing slurry to centrifugal force generated in a spinning vessel and causing a separation to occur due to a density difference between the solids and the liquid. Of all the available types of centrifuges, only two have proved suitable for deashing coal liquid extracts and have so far been widely tried by various investigators for this purpose with mixed success. These two types of centrifuges are the nozzle-disc and the solid-bowl centrifuges. A schematic diagram(4) of a nozzle-disc centrifuge is shown in Figure 2-1. Feed enters the unit from the top or bottom, and is distributed outwardly at the center of the rotating bowl, thus allowing the liquid to flow inward between a stack of closely-spaced (about 0.05 inches spacing) conical discs. The concentrated solids slurry is discharged through nozzles at the periphery of the rotating bowl. In general, the liquid clarification capabilities of the centrifuges are improved by increasing rotational speed, increasing the diameter of the discs, and decreasing the spacing between the discs. Solids concentration in the feed and in the underflow can also affect performance. This process has been tried with mixed success for deashing of coal liquid extracts by various investigators.

Two types of solid-bowl-scroll discharge centrifuges were tested for removal and concentration of solids from liquefied coal. These types, one horizontal and one vertical, are illustrated

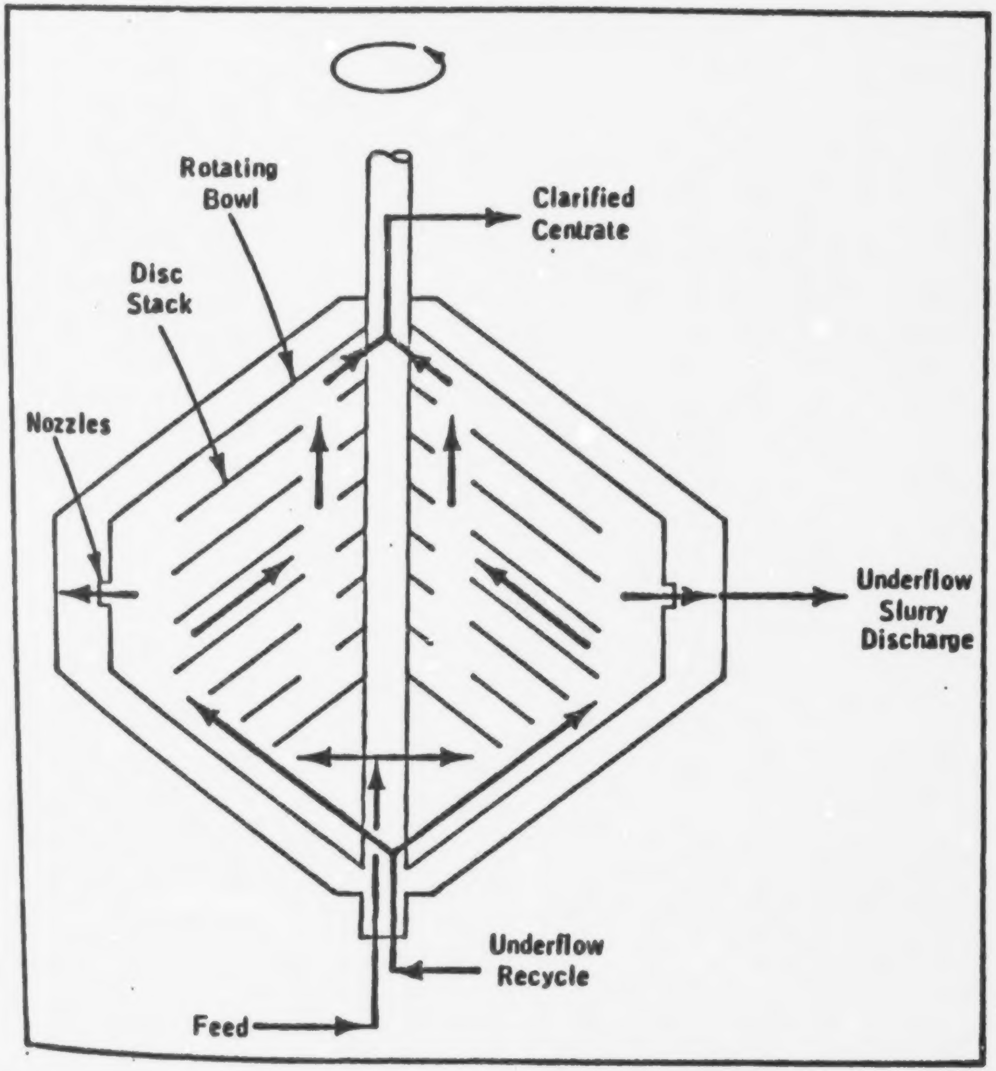
Rotating Bowl
Clarified Centrate
Disc Stack
Nozzles
Underflow Slurry Discharge
Underflow Recycle
Feed

FIGURE 2-1

DIAGRAM OF NOZZLE-DISC CENTRIFUGE

Source: Reference 4.

schematically in Figures 2-2 and 2-3 respectively. Slurry feed enters these centrifuges near one end of the rotating bowl; solids are forced to the outside walls and pushed along by a scroll conveyor for continuous discharge. The liquid flows inward and is skimmed at the opposite end.

For solid-bowl centrifuges, it is not generally possible to predict the solid removal efficiency on a theoretical basis alone, because of flow disturbances and non-ideal properties of liquid-solid systems. A semi-empirical approach, known as the Sigma concept, has therefore been developed for correlation/scaleup purposes. In this approach, the centrifuge geometry and rotational speed are used to calculate the Sigma value which characterizes such centrifuges. The Sigma value (Σ) represents the equivalent gravity settling tank area of a given centrifuge, and is estimated as the product of a mean sedimentation area and the equivalent gravitational force exerted by the centrifuge.(5)

Historically,(3) the Germans and British used centrifuges in their coal liquefaction plants of the 1940's, to remove solids which accumulated in the heavy-oil letdown product (HOLD). The solids ranged between 20 to 40 weight percent, and varied in size from 40 microns to submicrons. A diluting oil was added to the HOLD and the mnixture centrifuged. The result was not entirely satisfactory. The centrifuge concentrate was coked to recover the diluting oil.

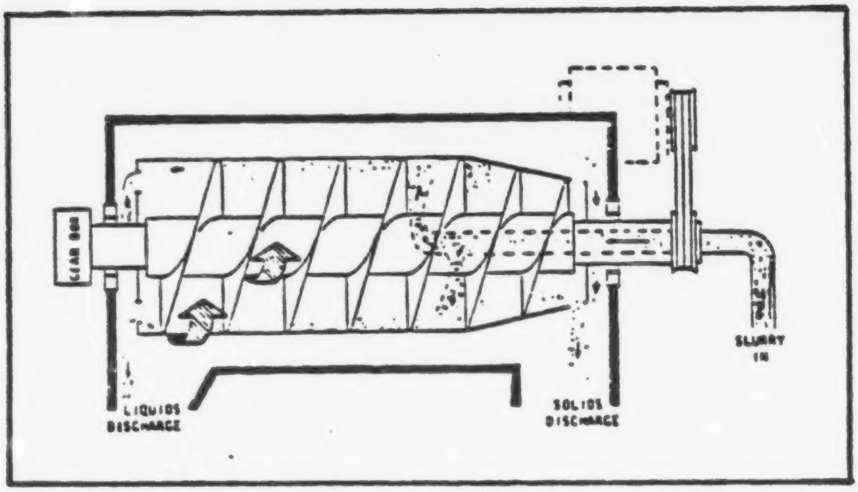
GEAR BOX
LIQUIDS DISCHARGE
SOLIDS DISCHARGE
SLURRY IN

FIGURE 2-2

SCHEMATIC OF HORIZONTOL SCROLL CENTRIFUGE

Source: Reference (4)

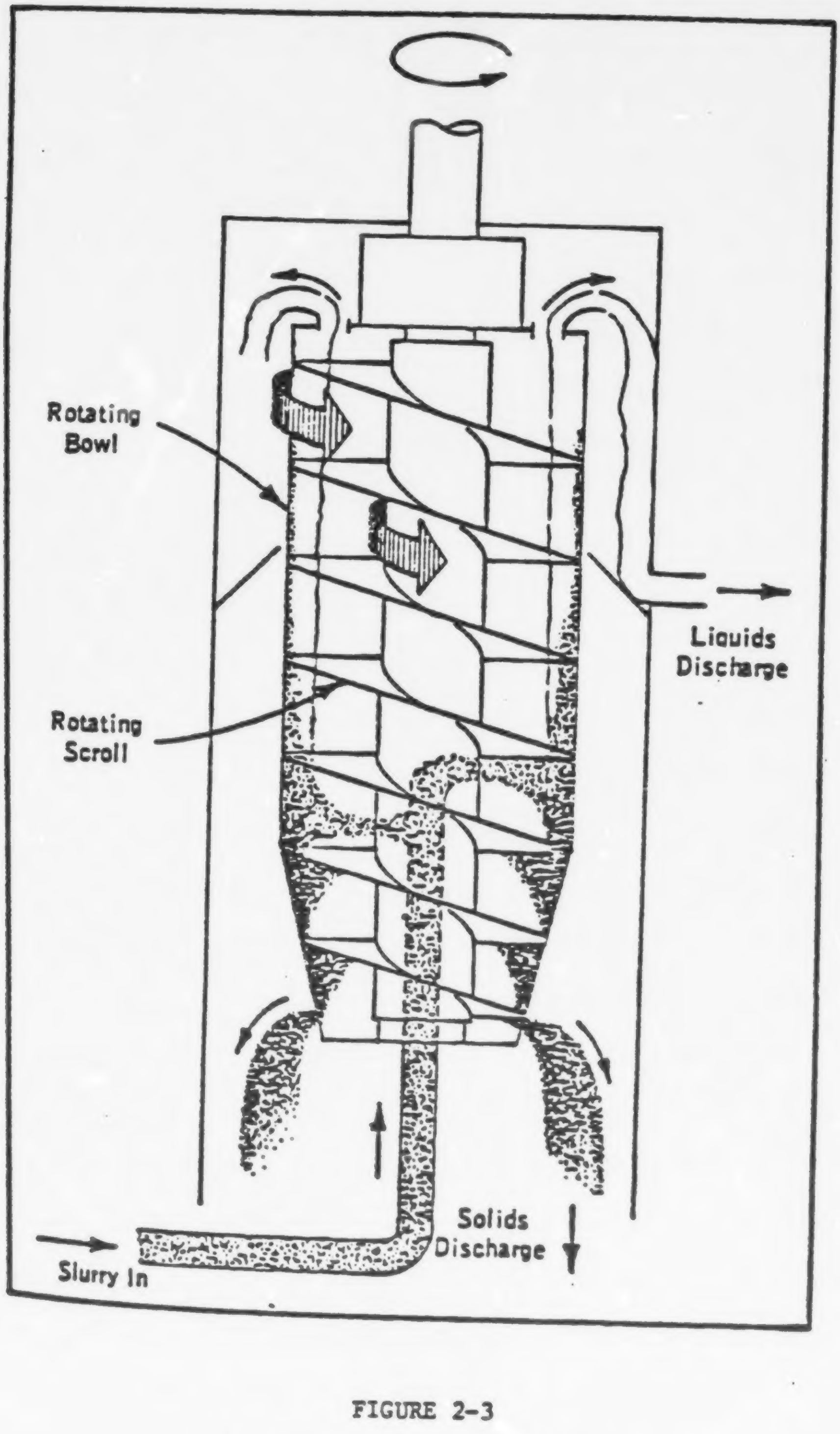
Rotating Bowl
Liquids Discharge
Rotating Scroll
Solids Discharge
Slurry In

FIGURE 2-3

DIAGRAM OF VERTICAL SOLID BOWL
SCROLL-DISCHARGE CENTRIFUGE

Source: Reference (4)

More recently, in development of the SRC process, Pittsburg and Midway[(3)] tested centrifugation of coal liquid extracts in both the bench-scale tests and the Process Development Plant (PDP). The bench-scale tests were promising, but liquid residence times greater than those feasible for commercial equipment were required to accomplish satisfactory levels of solids separations.

A disc-bowl type centrifuge (Sharples Model DHL Nozeljector) which was designed to operate at 8,000 rpm, 500°F, and 150 psig, was used for testing in the Pittsburg and Midway PDP.(3) Fifteen tests were performed with the centrifuge at temperatures ranging from 175° to 426°F. Up to 92 percent of the ash was removed by the centrifuge, but the lowest ash level in the centrate was only 0.1 to 0.2 percent at a feed rate of 100 lbs/hour. This did not compare favorably with an ash level of 0.03 to 0.04 in the filter effluent under the same operating conditions.

In the Solvent Refined Coal (SRC) pilot plant at Fort Lewis, Washington, where centrifugation has been tested at relatively low temperatures (200°F), the solid separation was poor, compared to filtration.(3) The centrifuge system consisted of three machines—two of the nozzle-disc type, and one of the solid-bowl type. The first nozzle-disc centrifuge was used to accomplish the key separation and to provide the solids-free coal solution. The second nozzle-disc centrifuge was used to wash the solids from the first machine. The solid bowl centrifuge was used to wash the solids from the second machine and provide an essentially dry stream of solids.

The SRC pilot plant at Wilsonville, Alabama, also conducted centrifugation tests, using a Sharples P-850 solid-bowl centrifuge as the primary solid separation device.(3) The throughput was very low when satisfactory deashing was accomplished. Plugging of centrifuges was frequent and required extensive maintenance.(3)

Other processes for which centrifugation was considered include Synthoil, CONSOL, H-Coal, and Exxon Donor Solvents (EDS). In the Synthoil process, the products from four hours of plant operation amounting to 90 to 95 pounds with an ash content of 6 to 8 percent were batch centrifuged at 80°C in a four-inch vertical Sharples supercentrifuge.(6) Thirty minutes were adequate for removing most of the solids and resulting in a centrate with about 2 percent ash. Longer run time resulted in negligible additional amounts of solids removal. Further tests using a Sharples Super-D-Canter solid-bowl centrifuge at 200°F indicated that centrifugation was less effective than rotary drum filtration for solids removal and sulfur reduction.

For the CONSOL process, a six-inch horizontal Bird continuous solid-bowl centrifuge developed problems with the seal.(6) These problems were attributed to the fluctuating pressure which results from the alternating pressurizing and venting of the lock hopper in discharging the recovered solid cake. Because of the seal difficulties, no reliable data were obtained.

Three different types of centrifuges were tested in the H-Coal process. These include a horizontal scroll type Sharples P-660 centrifuge, a vertical scroll type Sharples P-850 centrifuge, and a nozzle-disc (Dorr-Oliver Merco PC-9) centrifuge.(7) The vertical scroll unit was found to be more suitable for this application than the horizontal, for a variety of reasons. It was less susceptible to problems caused by load imbalance and had only one seal compared to three seals for the horizontal unit. Also, the vertical unit could run at 350°F and 150 psig compared with 250°F and atmospheric pressure for the horizontal unit. In terms of ash removal, however, the performance of the vertical unit was similar to that of the horizontal. Both units resulted in an ash removal efficiency ranging from 58 to 97 percent depending on the operating conditions.

The nozzle-disc unit was used as a second-stage clarifier for the overflow product from the scroll-type centrifuge. In such an arrangement, the nozzle-disc centrifuge resulted in a low-ash centrate with 0.1-percent ash. The conclusions of these tests were that solids separation by centrifugation methods will require a two-stage operation regardless of the type of centrifuge used.(7)

The extensive EDS pilot-scale evaluation of coal extract deashing techniques(4) involved testing of nozzle-disc centrifuges or solid-bowl-scroll centrifuges, and the use of antisolvent agglomeration to enhance the performance of the centrifuges.

A schematic flow sheet of the EDS test unit is shown in Figure 2-4. The feed to the centrifuge was a coal liquefaction pilot plant (CLPP-1) product generated from -100-mesh Illinois coal and hydrogenated creosote oil in the ratio of one part coal to two parts solvent, at 730°F and 350 psig.(4) The concentration of solids in the feed to the nozzle-disc centrifuge was 6 to 7 weight percent, including about 3 weight percent ash, with a median particle size of 7 microns.(4) Performance data from a typical series of runs through the nozzle-disc centrifuge at 420°F to 430°F are shown in Figure 2-5. This figure indicates that clarity improves linearly with decreasing throughput. At 30 lbs/minute, which is a typical commercial flow rate for this centrifuge, the centrate contains about 0.5-percent ash. The concentration of solids in the underflow stream varied for different runs but was generally about 15 to 18 percent.(4) The conclusion is that, while operation of this type of centrifuge appeared to be better suited than precoat filters for liquefied coal processing, it was clear that the capacity of the centrifuge is significantly reduced if the ash in the clarified product is required to be as low as 0.1 percent or less.(4)

To increase the capacity and separation efficiency of centrifuges, the addition of an antisolvent to help agglomerate the fine particles in the coal extract feed to the centrifuge was explored. In one experiment, the addition of 20-percent naphtha as an antisolvent reduced the ash content of the centrate from 0.4 to

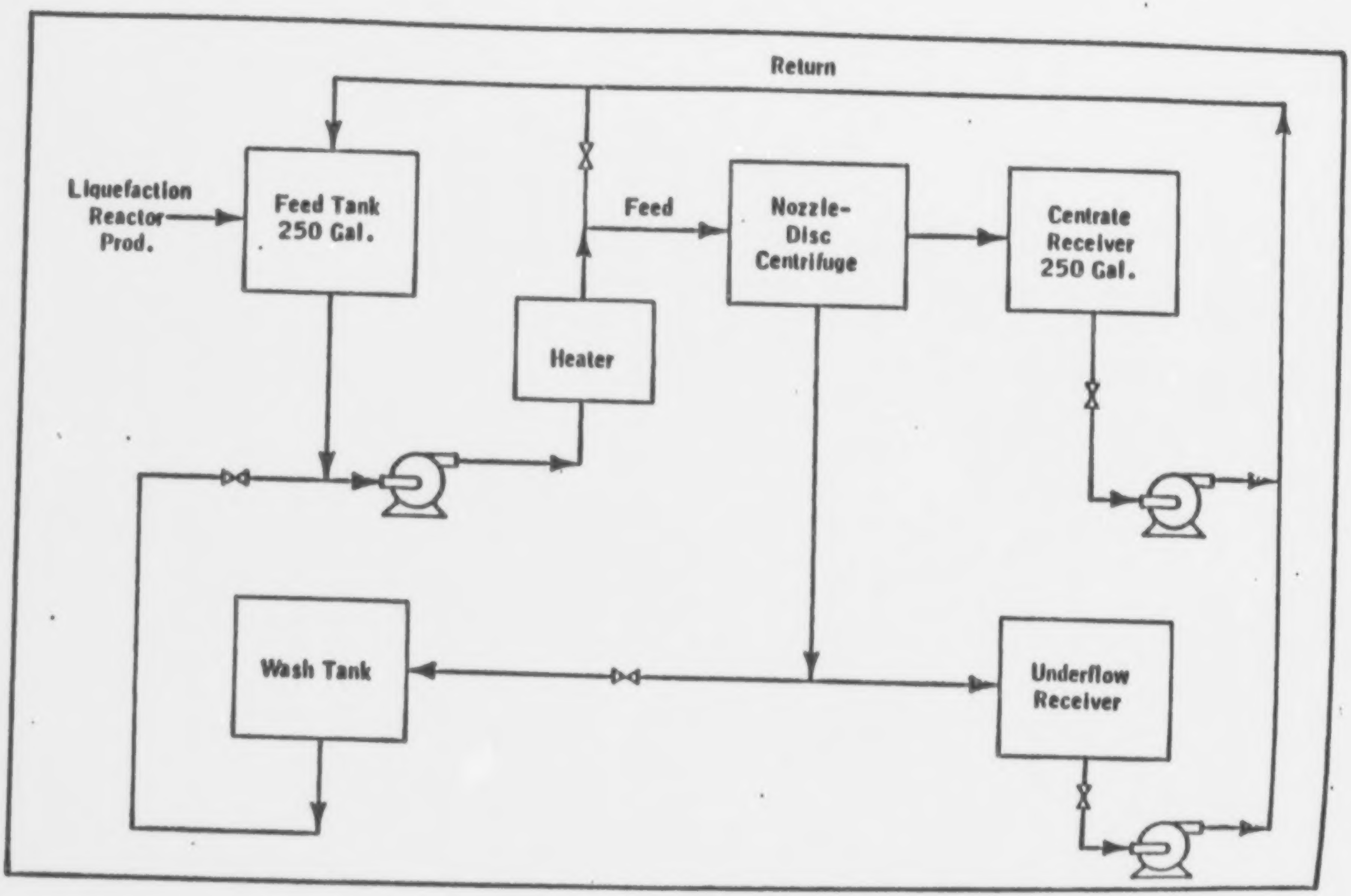
Return
Liquefaction
Reactor
Prod.
Feed Tank
250 Gal.
Feed
Heater
Nozzle-
Disc
Centrifuge
Centrate
Receiver
250 Gal.
Wash Tank
Underflow
Receiver

FIGURE 2-4

SCHEMATIC OF CENTRIFUGE "PAD" SYSTEM OF EDS

Source: Reference 4.

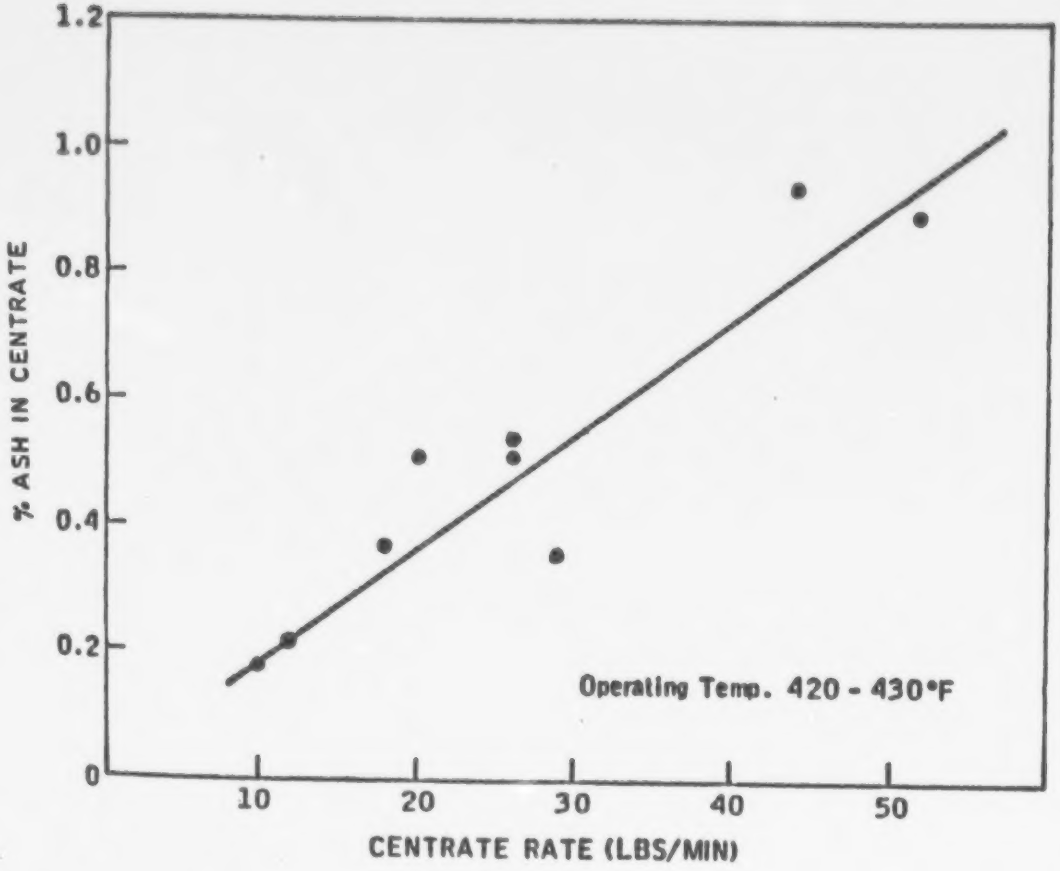

FIGURE 2-5

NOZZLE-DISC CENTRIFUGE PERFORMANCE

Source: Reference (4).

0.1 percent at approximately the same output rate and temperature.(4)

Similar tests were carried out in the EDS pilot-scale test unit on both the horizontal and vertical solid-bowl scroll centrifuges to obtain data on ash removal efficiency as a function of throughput (Q), the Sigma value of the centrifuge, and the operating conditions. Such ash removal data are depicted in Figure 2-6. At a Q/Σ value of 2×10^{-4} cm/sec (which represents the approximate value estimated for a commercial solid-bowl centrifuge operation), 38 percent of the original ash in the feed is not removed and remains in the centrate. Even if the throughput Q were halved, 26 percent of the original ash would remain in the centrate. The concentration of solids in the underflow generally ranged from 40 to 45 percent.

In general, the test data showed that the solid-bowl-scroll centrifuges could be suitable for concentrating solids, but for high centrate clarity, unreasonably low flow rates were required. The combination of a solid-bowl centrifuge followed by a nozzle-disc centrifuge appeared best suited to achieve both clarity of centrate and solids concentration. The use of solid-bowl ahead of the nozzle-disc provides additional protection of the nozzles from oversize particles.(4)

In all cases, the addition of an antisolvent greatly improved centrifuge operation.(4) This conclusion has been recently confirmed by the Hokkaido Technological Research Laboratory(8) of

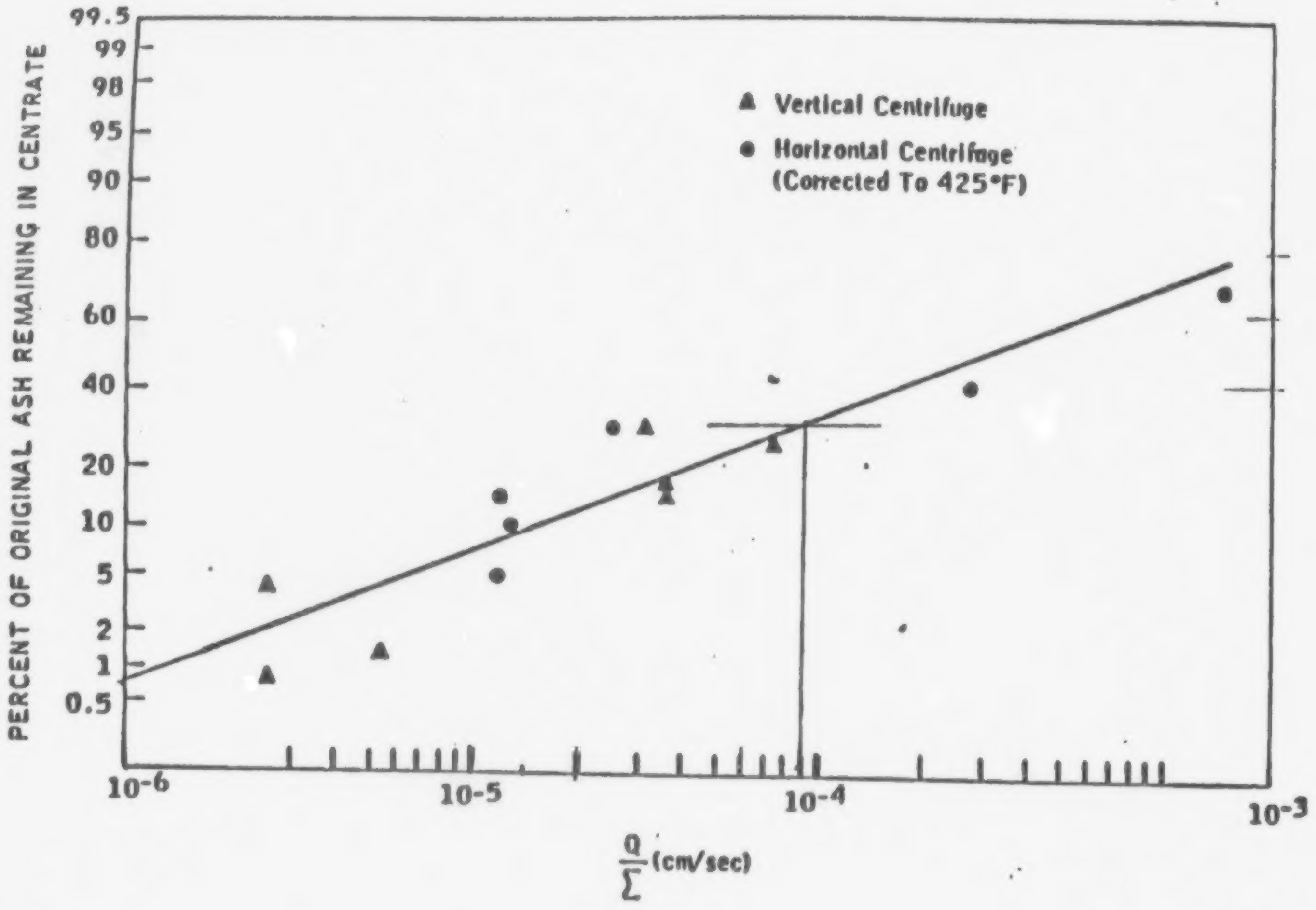

FIGURE 2-6

PERFORMANCE DATA ON SOLID BOWL SCROLL - DISCHARGE CENTRIFUGES

Source: Reference (4).

Japan, in a 3-ton/day, centrifugal solid/liquid separation plant, using an Australian subbituminous coal extract. Typical operating conditions for the plant are shown in Table 2-2. The hydrogen donor solvent used in the coal liquefaction (extraction) stage is petroleum-based with 44 weight percent of alkyltetraline and a boiling range of 242°C to 317°C.

Two experimental sets of centrifugal deashing tests of Australian coal extract were conducted. In both sets, two vertical scroll-tupe Sharples centrifuges were used. The first centrifugal unit was used to separate the coal extract feed into an overflow stream of fairly solids-free coal extract and an ash concentrate underflow stream which constitutes the feed to the second centrifugal unit. The second centrifugal solids separation unit recovers most of the clean coal extract retained in the underflow stream of the first centrifuge, leaving an almost dry ash concentrate.

It should be noted, however, that in the first experimental set, two conventional Sharples centrifuges were used with the addition of varying amounts of antisolvent (up to 40 weight percent of the coal extract feed to the first centrifuge) to help agglomerate the fine ash particles and enhance the centrifugal solids separation and ash removal process. The addition of 20 weight percent of antisolvent resulted in a solids separation and ash removal of about 99 weight percent, yielding a solids-free coal extract (i.e., SRC) product with less than 0.2 weight percent

TABLE 2-2

OPERATING CONDITIONS OF THE JAPANESE COAL EXTRACT DEASHING EXPERIMENT

Extractive Conditions					Conditions of Centrifugation					
Coal Concen. Wt.%	Donor Quantity Ratio to Coal	Temp. °C	Pressure kg/cm^2G	Time Min.	Rate of Rotat. rpm	Gravity	Δr** rpm	Feed Temp. °C	Pressure kg/cm^2G	Addition of Anti-Solvent* Wt.%
29	1,0	440	40	30	6000	3100	50	250-300	7-8	0-40

*Ratio to the extractive slurry

**Δr differential revolutions/minute between the solid bowl and the scroll.

Source: Reference (8).

ash.(8) Further addition of antisolvent was found to be ineffective.(8)

In the second set of experiments, a number of proprietary modifications were made on the conventional Sharples vertical scroll-type centrifuges to improve the clarity of their overflow stream without the addition of any antisolvent. The performance of the modified Sharples centrifuge was not significantly different from that of the convential Sharples centrifugal separation unit when 20 weight percent of an antisolvent was used. It appears that the modified centrifugal separation unit has a lower recovery efficiency of preasphaltene in comparison to the conventional Sharples model. This allows an increased sedimentation of preasphaltene which enhances the flocculation of the fine solids. As a result, the modified centrifugal unit can remove the particles which are near the separation limit for the conventional Sharples vertical scroll-type centrifuge, even without the addition of an antisolvent.

In a recent experimental study conducted by the Pittsburgh Energy Technology Center (PETC)(9), the application of the continuous solid-bowl-scroll centrifuge to the removal of solids from coal liquefaction products was investigated. The centrifuge removed from 23 to 88 percent of the input ash and from 8 to 73 percent of the input organic benzene insolubles, depending upon operating conditions. Organic solids removal was poorer and more erratic than ash removal because of the lesser density differences and greater size variability of the organics.

All previous studies on the evaluation of centrifugal deashing of coal extract seem to indicate that a centrifuge cannot currently achieve as sharp a solid separation as a filter. The major limitations of centrifuges are:(3)

- Their incapability of removing the very fine particles (2 to 3 microns) usually found in liquefied coal slurries.
- Their inability to achieve an acceptable low ash level in the centrate at a reasonable flow rate. In removing solids, 1 to 5 percent of the solids feed to the machine would usually remain in the discharged liquid.
- Their need for a better solids discharge system.
- Their relatively high operating energy requirement.
- The currently available commercial designs are not suitable to handle the operating conditions of 150 to 200 psig and 600° to 700°F, which are normally required for coal liquefaction processes. The state-of-the-art machines can handle temperatures of up to 460°F and pressures not higher than 150 psig.
- Their high maintenance requirement.
- The low environmentally acceptable limits on the noise level of high speed rotating machinery.
- Lack of adequate scale-up basis to commercial size machines.
- The small size of commercial units which is imposed by the structural stress limits on the available material of construction used to build different centrifuges.

Solvent Deashing Techniques

Because of its promising potential, solvent deashing is the basis for some newly introduced deashing techniques. These techniques include Lummus antisolvent deashing (ASDA) and the Kerr-McGee critical solvent (CSD) deashing techniques.

The principle of antisolvent deashing involves the addition of a foreign solvent to the coal extract, which initiates the precipitation of a fraction of the preasphaltenic material. This precipitation results in an agglomeration and growth of the particulate matter which can settle out of the coal extract under gravity. The antisolvent also acts to decrease the viscosity of the coal extract so that the settling can occur in a relatively rapid time frame. Deashing is accomplished by withdrawing the solids-laden lower phase (i.e., ash-concentrate) from a gravity settler vessel. This is the mode of operation of the Lummus ASDA which has been successfully tested for several years on a continuous process development unit (PDU) scale with a daily coal feed rate of up to 700 pounds of bituminous and subbituminous coal. In many cases, ash removal efficiencies of more than 98 percent with recoveries not less than 86 percent of heavy, nondistillable liquefied coal product have been attained. The overflow from the settler contains about 0.1 weight percent ash. The underflow contains approximately 40 to 45 weight percent solids.(10)

Kerr-McGee has developed a solvent deashing process which operates at conditions close to the critical temperature and pressure of the selected solvent. This critical solvent deashing (CSD) process is basically a selective solvent extraction process.

The feed to the unit is a stripped residual (+850°F) material from vacuum distillation of the product from the first-stage coal liquefaction reactor. Extraction of this material occurs in the CSD first-stage settler at temperature and pressure conditions which increase the solvent power of the critical solvent. Under these conditions, a heavy phase, consisting of unextracted coal liquids and particles of IOM and ash, is separated by gravity settling from the deashed coal solution phase. Subsequent vaporization and recovery of the deashing solvent from the heavy phase leaves a free-flowing, fine powder known as ash concentrate. This dry ash concentrate, produced by essentially spray drying the heavy phase, still contains a quantity of rejected coal solution, but less than from antisolvent deashing, since the requirement to pump the underflow from the settler is obviated in the case of a free-flowing solids reject stream. The Kerr-McGee CSD process has been successfully demonstrated for many years on pilot scale with a daily coal feed rate of six tons, at Wilsonville, Alabama.

Both the Kerr-McGee and Lummus deashing processes result in a loss of extracted material ranging from 8 to 14 weight percent of the coal (moisture-free). In addition, comparative economic analyses of these processes have shown that the deashing unit represents a significant capital cost that amounts to more than 11 percent of the cost of a full-scale ITSL plant.(11) Ample opportunities exist to improve product recovery yields and reduce capital and operating costs.

The CENTRAX Separator

The Canadian Sandwell Beak Research Group (SBRG) CENTRAX separator is a recent attempt to overcome the shortcomings of existing centrifugal separators.(12) The CENTRAX separator is basically a combination of a newly-designed prototype of a multistage solid-bowl centrifugal solid separation and extraction machine. The first centrifugal stage is used for solid separation and to provide a reasonably solid-free coal solution. The remaining centrifugal stages are used to extract and wash the solids separated in the first stage and provide an essentially dry stream of solids while maximizing liquid recovery.

This CENTRAX separator is claimed to have achieved higher solid removal rates than those demonstrated in the Lummus ASDA and to have allowed an almost complete recovery of liquefied coal product (about 99 percent) with minimum energy requirement of about 1.4 percent of the heating value of the feed coal. The energy rejected by the CENTRAX separator in the ash concentrate, expressed as a percentage of the energy in the feed coal, is claimed to be between 10 and 18 percent compared to an energy rejection of 20 to 25 percent for the Kerr-McGee CSD process when applied to the same feedstock.

It should be noted, however, that the CENTRAX process is still in its very early concept development stage. To date, the proof of concept has only been tested on a batch laboratory bench scale. The

SBRG are now attempting to confirm their laboratory results on a prototype CENTRAX separator at a coal extract feed rate of up to four gallons per minute. If successful, the CENTRAX separator may eventually represent a significant improvement on the current state-of-the-art of coal liquid deashing technology.

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